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CONSUMER PROTECTION AND ENVIRONMENTAL HEALTH SERVICE
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION
DIVISION OF PROCESS CONTROL ENGINEERING
DIVISION OF ECONOMIC EFFECTS RESEARCH

CONTROL OF ATMOSPHERIC EMISSIONS IN THE WOOD PULPING INDUSTRY

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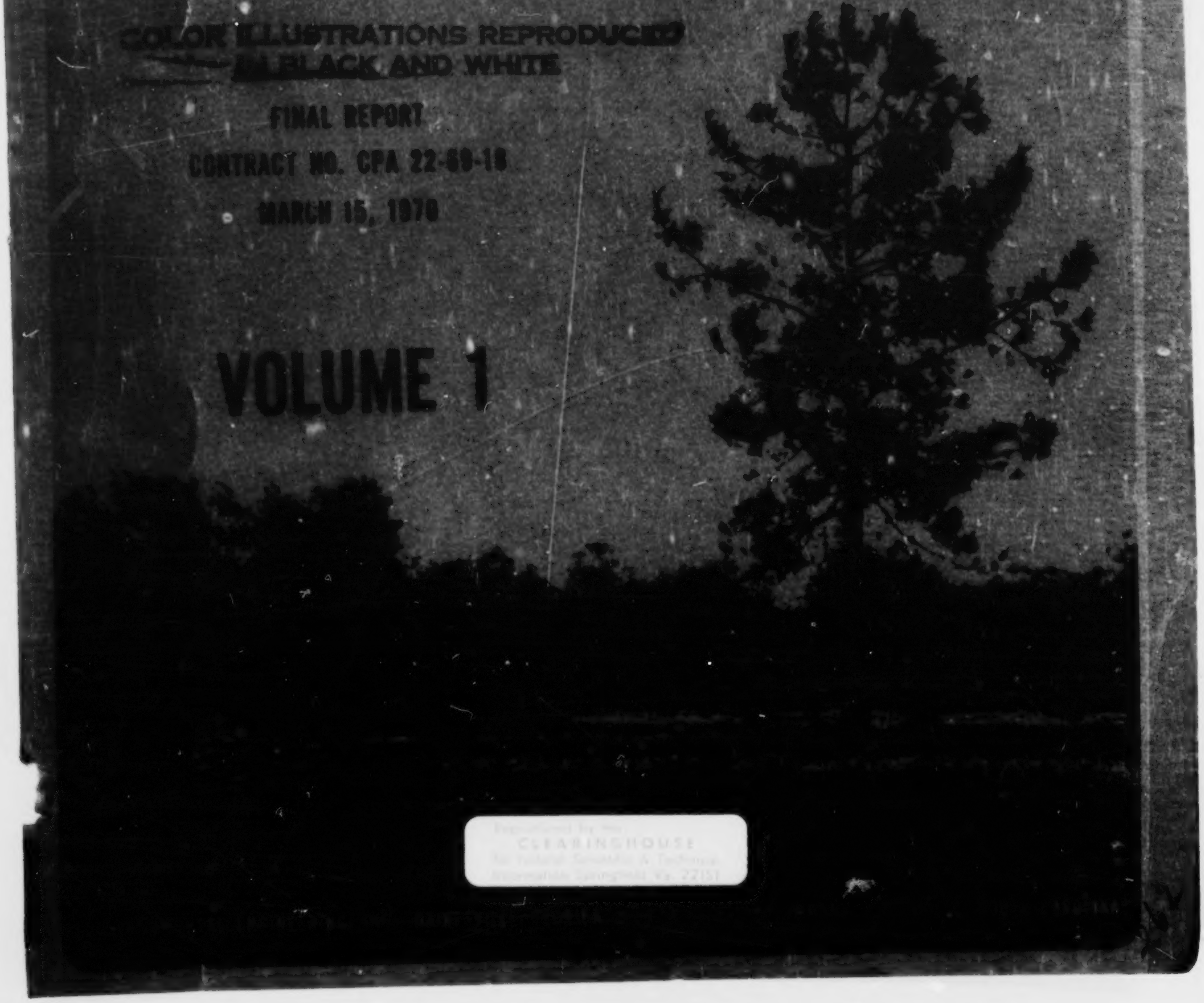
FINAL REPORT

CONTRACT NO. CPA 22-69-18

MARCH 15, 1970

VOLUME 1



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ERRATA SHEET

△ SHOULD BE □ IN REFERENCE FIGURE 2-2 ENTITLED "REGIONAL DISTRIBUTION OF SULFITE AND NSSC PULP MILLS IN THE U.S."

SULFITE MILL SHOWN IN WESTERN NORTH CAROLINA SHOULD BE NSSC.

DEPARTMENT OF HEALTH, EDUCATION, AND WELFARE
CONSUMER PROTECTION AND ENVIRONMENTAL HEALTH SERVICE
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FINAL REPORT
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VOLUME 1

by

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Principal Investigator

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ENVIRONMENTAL ENGINEERING, INC., GAINESVILLE, FLORIDA
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SYSTEMS ANALYSIS STUDY OF EMISSIONS CONTROL IN THE WOOD PULP INDUSTRY
CONSULTANTS Professor Donald F. Adams Poly Con, Inc.

ENVIRONMENTAL ENGINEERING, INC.
J. E. SIRRINE COMPANY
Gainesville, Florida • Greenville, South Carolina

15 March 1970

Mr. W. Gene Tucker
Division of Process Control Engineering
National Air Pollution Control Administration
5710 Wooster Pike
Cincinnati, Ohio 45227

Mr. F. L. Bunyard
Division of Economic Effects Research
National Air Pollution Control Administration
1033 Wade Avenue
Raleigh, North Carolina 27605

Re: Final Report, Contract No. CPA 22-69-18

Gentlemen:

Fulfilling the requirements of Contract No CPA-22-69-18, we have prepared for NAPCA 200 copies of the final report entitled "Control of Atmospheric Emissions in the Wood Pulping Industry." For ease in handling, the report has been bound in three volumes. Each chapter is headed by a separator sheet which contains the complete table of contents for that chapter, and in addition each volume contains a table of contents for all three volumes.

In accordance with your letter of 21 February 1970, we are shipping 125 copies of the complete report to Cincinnati and 75 copies to Raleigh.

It has been a pleasure serving NAPCA and the cause of cleaner air in fulfilling this contract.

Sincerely yours,

E. R. Hendrickson, Ph.D., P.E.
Principal Investigator

ERH/gea

2324 S. W. 34th Street • Gainesville, Florida 32601 • 904/372-3318

ABSTRACT

The basic objectives of this study were to make a comprehensive and systematic evaluation of the technical and economic problems involved in the control of airborne emissions, especially particulates and gaseous sulfur compounds from the chemical wood pulping industry; and to determine the technological gaps that need to be filled by accelerated research and development.

Included in the scope of the work were major variations of the kraft, sulfite, and semichemical pulping processes; the nature and sources of emissions from each process; a review of control hardware capabilities, efficiencies, and costs; a review of source and ambient air sampling and analysis techniques; and an evaluation of the overall economic impacts of air quality improvement in the industry.

It is felt that several major gaps in technology have been identified which will need to be filled before any further great steps in progress can be taken. Brief statements of these needed areas of research of highest priority are as follows:

1. Develop and standardize methods and instruments for monitoring emissions and ambient air.
2. Assess the effect of operating variables on emissions from the kraft pulping and recovery systems.
3. Develop and standardize organoleptic techniques for determinations of process emissions and evaluation of ambient air quality.
4. Investigate new pulping methods which eliminate the use of sulfur.
5. Define the mechanisms, with emphasis on transport processes and emission interactions, which will relate emission limitations to ambient air objectives.
6. Evaluate emissions from sources in sulfite and NSSC mills and determine operating variables which affect emissions.

7. Investigate adsorption and absorption of odorous gases and reuse of the collected material in process.

8. Determine whether TRS is an effective measure of the acceptability of odorous emissions from kraft mills or must the compounds be identified more definitively.

These brief statements of needs are defined more completely and specific projects identified.

ACKNOWLEDGMENTS

In a project of this magnitude, there are many people other than those designated as the authors who have made significant contributions to the Final Report.

In Environmental Engineering, Inc., Mr. Kent Withington assembled the necessary information and prepared the first drafts of the chapter on emissions. Dr. David T. Knuth and Mr. John Dollar assisted with the literature search on on-going research. At Reynolds, Smith and Hills, Mr. Lamar Russell was originally the project engineer, but was succeeded by Mr. Leroy Doughty about the mid-point in the project. Mr. Doughty was assisted by Mr. Robert Clark in collecting the information about current expenditures by the industry. Mr. Robert Clark, Mr. Forrest Dryden, and Mr. Malcolm Steeves investigated sulfur recovery from power boilers and prepared the drafts of that chapter. All RSH and EEI personnel participated in the selection of a plan for projecting investment and operating costs. Mr. Doughty of RSH was primarily responsible for this section working with Dr. James Heaney of EEI who was responsible for the modeling.

From the Sirrine organization, Mr. Robert Farrell prepared the power plant energy balances assisted by Mr. H. J. Steigler. Mr. Carlton Ranew and Mr. Peter Gombola prepared the flow diagrams assisted by Mr. J. Don Lee, Mr. S. L. McCluskey, and Mr. J. D. Rushton. Mr. Wells Meakin conducted the survey and prepared the preliminary drafts on which Chapter 2 is based. Contributions to Chapters 5, 6, and 7 were made by the EEI personnel cited above and the following Sirrine personnel: Mr. J. H. Bringhurst, Mr. W. L. Carpenter, Mr. M. C. Freeland, Mr. P. P. Gombola, Mr. E. C. Hartney, Mr. C. E. Hatch, Jr., Mr. J. Don Lee, and Mr. R. C. Ranew, Mr. J. E. Roberson, Mr. J. D. Rushton, Mr. H. A. Stokes, Mr. J. W. Stubblefield, Jr., and Mr. D. B. Wilson.

Consultants to the contractors included Poly Con, Incorporated, of Ramsey, New Jersey. Mr. Jorgen Hedenhag and Mr. Samuel Jacobson prepared much of the background information on control equipment used in Chapters 5 and 6 and the basic cost data on control equipment used throughout the report.

Another consultant was Professor Donald Adams of Washington State University who prepared the material used in the chapter on sampling and analysis.

The guidance and assistance provided by the two project officers, Mr. W. Gene Tucker of the Division of Process Control Engineering, and Mr. Frank Bunyard of the Division of Economic Effects Research, National

Air Pollution Control Administration, is gratefully acknowledged. Mr. Tucker and Mr. Bunyard were able to furnish information which was not readily available from other sources. Both were of special help in polishing up the many drafts which led to the Final Report.

Liaison was maintained throughout the project with a committee drawn from the chemical wood pulping industry. These gentlemen contributed information to the contractors including otherwise unavailable cost data. Members of the pulp industry Liaison Committee to whom great thanks are due include Dr. Herman Amberg of Crown-Zellerbach; Mr. Richard Billings of Kimberly-Clark; Mr. Russell Blosser and Dr. Isaiah Gellman of National Council for Air and Stream Improvement; Dr. Loren Forman, Scott Paper Company (Dr. Nicholas Lardieri, alternate); Mr. Matthew Gould of Georgia Pacific; Dr. Glenn Kimble of Union Camp; Mr. G. J. Kneeland, St. Regis; Mr. George Marsh of U. S. Plywood-Champion; Mr. John McClintock of Weyerhaeuser; Dr. Samuel McKibbins of Continental Can; Mr. George Rand of International; Mr. J. T. Walker of Westvaco (Mr. Bill Wassmer, alternate); Mr. Peter Wrist of Mead (Mr. Virgil Minch, alternate).

The typing of the many drafts and the plates for the Final Report was done by Mrs. Peggy Bowman, Mrs. Mary Ann Hester, Mrs. Lala Scouten, and Miss Ann Smith.

PREFACE

This report was prepared for the National Air Pollution Control Administration to assist in carrying out their responsibilities under the Air Quality Act of 1967. Consideration was given mainly to the needs of the Division of Process Control Engineering and the Division of Economic Effects Research. It is probable that the information also will be used by others including persons in the chemical wood pulping industry.

The intention was to provide a report which essentially would specify the present status of emissions control in the industry, indicate what additional progress could be expected by application of existing or nearly-developed technology, and define areas of research and development necessary for further advances in the future.

As much background explanatory material as possible was provided. It is not necessary to be intimately familiar with the technical aspects and economics of the industry. However, use of the report presupposes technical knowledge of the processes used by the industry and an appreciation of emission control technology.

It is important that the report be considered in its entirety. It was impossible adequately to qualify all discussions and conclusions at the place the information appears in the report. Thus, erroneous conclusions may be drawn by taking material out of context without a proper understanding of the background.

Costs used in the engineering estimates and calculations were based on the price of supplies and equipment as of January 1969. Statistical data on the industry were verified as of December 1968.

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CHAPTER 1

INTRODUCTION

1.1 THE AIR QUALITY ACT OF 1967

The Air Quality Act of 1967 builds upon the basic precepts of the Clean Air Act of 1963 to develop a technically sound and rational plan for improvement of air quality through application of air quality criteria and detailed control technology. This would be accomplished on a regional basis by a cooperative effort on the part of the states and the Federal government. Key items in the abatement program would include designation of air quality control regions, publication of air quality criteria, publication of information on available control technology, development of air quality standards, and development of implementation plans to meet the standards.

The basic responsibility of the National Air Pollution Control Administration for research and development into the causes, effects, extent, prevention, and control of air pollution was expanded. This was necessary to provide an improved technological basis for the total program. Research and the identification of needed research were recognized as the key to effective air quality improvement. In addition, Congress expressed concern regarding the economic impact of implementing the legislation. They directed that specific economic studies be undertaken and a special report made on the costs to all segments of the economy of carrying out the provisions of the law.

The Air Quality Act retained many of the provisions of the Clean Air Act and, in addition, has many important provisions not reported here. The revisions described above, however, are the major ones which have a bearing on the need for this study.

1.2 SPECIAL INDUSTRY STUDIES

Some of the largest manufacturing industries have some of the most complex air quality control problems. Because of their technical orientation, the industries have been successful in developing new technology for solving their problems. In addition, the technology which they have developed may be applicable to control of other sources. The Air Quality Act imposed on NAPCA the responsibility of becoming familiar with the technology of air pollution control, developing control technology documents to accompany air quality criteria documents, identifying areas of research essential to progress in carrying out the provisions of the act, and assessing the economic impact of implementing the legislation. To carry out some of these responsibilities, the Division of Process Control Engineering and the Division of Economic Effects Research jointly planned and are conducting under contract, a series of systems analysis studies of emissions control in major industries. The study being reported is concerned with the control of atmospheric emissions in the chemical wood pulping industry.

1.3 OBJECTIVES OF THE STUDY

The purpose of this study was to make a comprehensive and systematic evaluation of the technical and economic problems involved in the control of airborne emissions, especially particulates and gaseous sulfur compounds from the wood pulping industry; and to determine the technological gaps that need to be filled by accelerated research and development. Included in the scope of the project were a consideration of major variations of the kraft, sulfite, and semichemical pulping processes; the nature and sources of emissions from each process; a review of source and ambient air sampling and analysis techniques; a review of control hardware capabilities, efficiencies, and costs; and an evaluation of the overall economic impacts of air quality improvement in this industry.

1.4 PROCEDURES FOR THE STUDY

The first phase of the project was concerned essentially with compiling and presenting both statistical and technical data about the chemical wood pulping industry. Information was gathered on the locations, types, and capacities of all chemical pulp mills in the United States. Projections of the data were made through 1980 in an attempt to present a picture of increases in capacity as well as changes in regional distribution and predominant pulping processes of the industry.

Surveys and critical reviews were made of ongoing research which is related to control of atmospheric emissions and of techniques used in the industry to monitor emissions. A major effort in this phase was devoted to defining typical pulping processes (flow diagrams) and preparing appropriate heat and materials balances including identification of emissions. Confirmation of the data developed during this phase was obtained by actual visits to selected operating mills in various parts of the country which had processes similar to the theoretical flow diagrams.

A second phase involved a feasibility analysis of emissions control technology in the industry. By appropriate modelling based on confirmed flow diagrams, the cost and effectiveness of emission control were evaluated for individual sources and for entire pulping processes. New foreign and domestic developments were considered in addition to current U. S. practices. The feasibility of sulfur recovery from utility boilers also was investigated.

Based upon all of the preceding work, gaps in technology were identified and recommendations made as to needed research and development efforts. The recommendations include, but are not limited to, the following areas:

1. The need for further development of, or clarification of effectiveness and cost of current control technology,
2. The necessity of developing new sampling and analytical techniques,
3. The necessity of research and development in new control technology, and
4. Modification of current pulping processes or the development of new processes that could lead to reduced atmospheric emissions.

The final phase of the work was an economic study of emission control in the pulping industry. A model was developed which utilizing the information gathered in previous phases made possible an estimation of the capital cost and annual operating and maintenance costs of pollution control equipment now in operation in the industry. A projection model was developed to provide the capability of projecting expenditures for achieving desired levels of emission control for a reasonable time into the future.

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THE CHEMICAL WOOD PULPING INDUSTRY

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CHAPTER 2

THE CHEMICAL WOOD PULPING INDUSTRY

SUMMARY

The pulp and paper industry is the ninth largest in the United States, accounting for nearly four percent of the value of all manufacturing. The per capita consumption of paper is expected to continue to rise from the late 1969 value of 550 pounds per year.

The United States and Canada produce more than 52 percent of the world's supply of pulp, with the U. S. in 1968 furnishing nearly 38 million short tons. Of this amount, 32 million tons were chemical pulp. Approximately 75 percent of this was produced by the kraft process, 9 percent by the sulfite, and 10 percent by neutral sulfite semichemical. The U. S. industry includes more than 360 pulp mills of all types, mechanical and chemical.

This study is concerned mainly with three types of chemical pulping processes; kraft, sulfite, and NSSC. The geographical distribution of the industry as of December 1968 by process and size is shown in the chapter by maps and in Appendix A by tables.

Projections have been made of chemical pulp production by process and region of the country through 1985. The production of soda and dissolving pulps is expected to remain reasonably constant. Sulfite pulp production will probably decrease slightly. It is anticipated that NSSC production will nearly double and kraft increase to approximately two and one-half times the 1968 figures.

By 1985, kraft and NSSC are projected to dominate chemical pulping in the U. S. with kraft accounting for 85 percent and NSSC 9 percent of total chemical pulp production.. The total production of chemical pulp is expected to slightly more than double over the 1968 figures.

Regional distribution of pulping capacity is expected to remain in the same relative proportions as it is today.

2.1 INTRODUCTION

The pulp and paper industry is the ninth largest manufacturing industry in the United States. This industry accounts for nearly four percent of the value of all manufacturing.

Although paper is one of the oldest manufacturing industries, it is an industry that is expanding faster than the general economy. Consumption of paper rises with increased affluence. The per capita consumption of paper in the U. S. is now about 550 pounds per year compared to about 412 pounds only ten years ago. There are no signs that per capita consumption is leveling off.

The pulp and paper industry is comprised of three distinct segments: (1) pulp, (2) primary paper and paperboard (cardboard, et cetera), and (3) converted paper and paperboard products.

Pulp--Most pulp is made by integrated companies and consumed captively without moving through the marketplace. About ten percent of the total pulp produced is, however, made by independent pulp producers without their own paper making facilities or by integrated companies producing surpluses for market. About three percent of all pulp produced is consumed outside the industry for such products as cellophane, rayon, cellulose esters and ethers, and their derivatives. Eighty percent of the pulp used for making paper comes from wood; about 20 percent of the pulp is made from waste paper or such fibers as cotton and bagasse.

Primary Paper and Paperboard--This segment of the industry produces paper, paperboard, and building paper and board. A portion of this production is sold directly to industrial users such as newspapers, book publishers, and printers, or to building and other users.

Converted Paper and Paperboard Products--About 70 percent of the primary paper production, however, is further processed by paper converters into such products as containers, bags, sanitary tissue products, and stationery.

Wood pulp is prepared either mechanically or chemically. In the mechanical processes (groundwood, defibrated and exploded) wood is shredded or separated by physical means. In the chemical processes (kraft, sulfite, NSSC, soda, and dissolving) wood is treated with chemical reagents which form soluble compounds with the noncellulosic materials, thus leaving residual cellulose. The NSSC process involves treating the wood first with a mild chemical and then mechanically separating the fibers.

Data are presented in Table 2-1 which show the number of mills and total mill capacities for the five chemical pulping processes. In Table 2-1, all capacities are based on air dried tons of pulp; annual capacities are based on operating at rated capacity for 350 days per year, allowing for normal maintenance and scheduled shutdowns. It is emphasized that these figures represent production capability and do not portray actual production data.

TABLE 2-1

SUMMARY - U.S.A.
CHEMICAL PULP MILL CAPACITIES (UNBLEACHED)

AS OF DECEMBER 31, 1968

Process	Number of Mills	Capacity* ADT/Day	Annual Capacity* Tons	1968 Production Tons
Kraft	116	87,808	30,733,000	24,300,000
Sulfite	43	10,875	3,799,500	2,500,000
NSSC	43	10,675	3,736,500	3,500,000
Dissolving	8	4,565	1,600,000	1,500,000
Soda	4	570	200,000	200,000
TOTALS	214	114,473	40,069,000	32,000,000

*These figures represent capacity and not actual production. ADT stands for air-dried tons of unbleached pulp per day; air-dried pulp contains 10 percent moisture.

Three of the chemical wood pulping processes display a potential to cause air pollution. These are the kraft, sulfite, and NSSC processes. Collectively, these three processes account for about 80 percent of the total wood pulp produced in the United States.

To define the air pollution problem posed by the chemical wood pulping industry, it is necessary to establish the geographical distribution of existing production capacity and to identify trends which might cause a redistribution of production capacity in the future.

2.2 ECONOMIC POSITION

The United States pulp and paper industry includes more than 360 pulp mills of all types, mechanical and chemical. Estimates for 1967 indicate 37 companies, each with pulp and paper sales at the manufacturer's level of at least $100 million, accounted for $10.24 billion in sales, or 49 percent of the industry's total of $20.88 billion.

Table 2-2 shows the wood pulp production in 1968 for the ten leading pulp producing nations of the world (1). The other 60 pulp producing nations individually produced less than one million tons of pulp and collectively produced 13,441,000 tons of pulp in 1968. From these data, it can be determined that the U. S. and Canada produced 52 percent of the world's wood pulp in 1968. This massive capacity, coupled with the contiguous features of the U. S. and Canada, place these countries in a leading position in terms of production.

It is reported (2) that North American industry is planning to build 65 new pulp mills in the early 1970's--39 in the U. S. and 26 in Canada. It seems reasonable to conclude, therefore, that the U. S. and Canada will remain the dominant nations in wood pulp production at least for the next two or three decades.

TABLE 2-2

PRODUCTION OF TEN LEADING PULP PRODUCING NATIONS - 1968

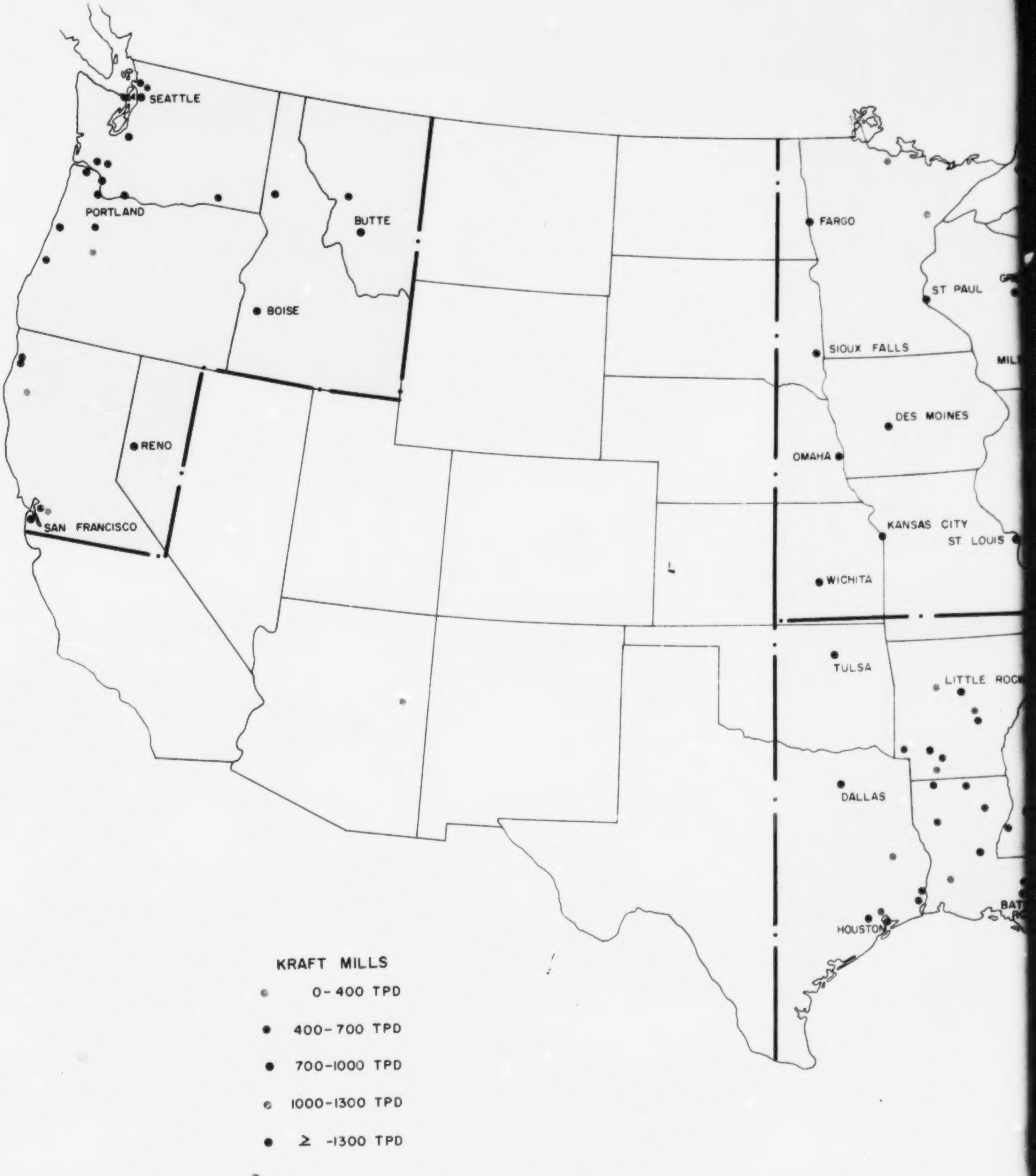
Nation	Million Short Tons
United States	37.89
Canada	16.40
Sweden	7.76
Japan	7.56
USSR	6.78
Finland	6.56
Mainland China	2.30
Norway	2.18
France	1.77
West Germany	1.73

Data taken from Pulp and Paper: 19th Annual World Review

2.3 PRESENT GEOGRAPHIC DISTRIBUTION

A compilation of data on current wood pulping practice in the U. S. was begun by searching available published reports, such as Lockwood's *Directory of the Paper and Allied Trades*, Post's *Pulp and Paper Directory*, and Southern Pulp and Paper Manufacturer's *Southern Mill Directory*. Information originally tabulated included plant location, owner, pulping process employed, rated mill capacity, type of wood pulped, and age of original mill. These data were brought up to date based on in-house information and available NAPCA - NCASI reports, and communications with mill managers.

The corrected data have been tabulated and are presented in Appendix A (Tables A-1 and A-2). Using these data as a base, two maps have been prepared to illustrate geographically the distribution of chemical wood pulping mills throughout the United States. These maps are presented here as Figures 2-1 and 2-2.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● 400-700 TPD
● 700-1000 TPD
● 1000-1300 TPD
● ≥ -1300 TPD

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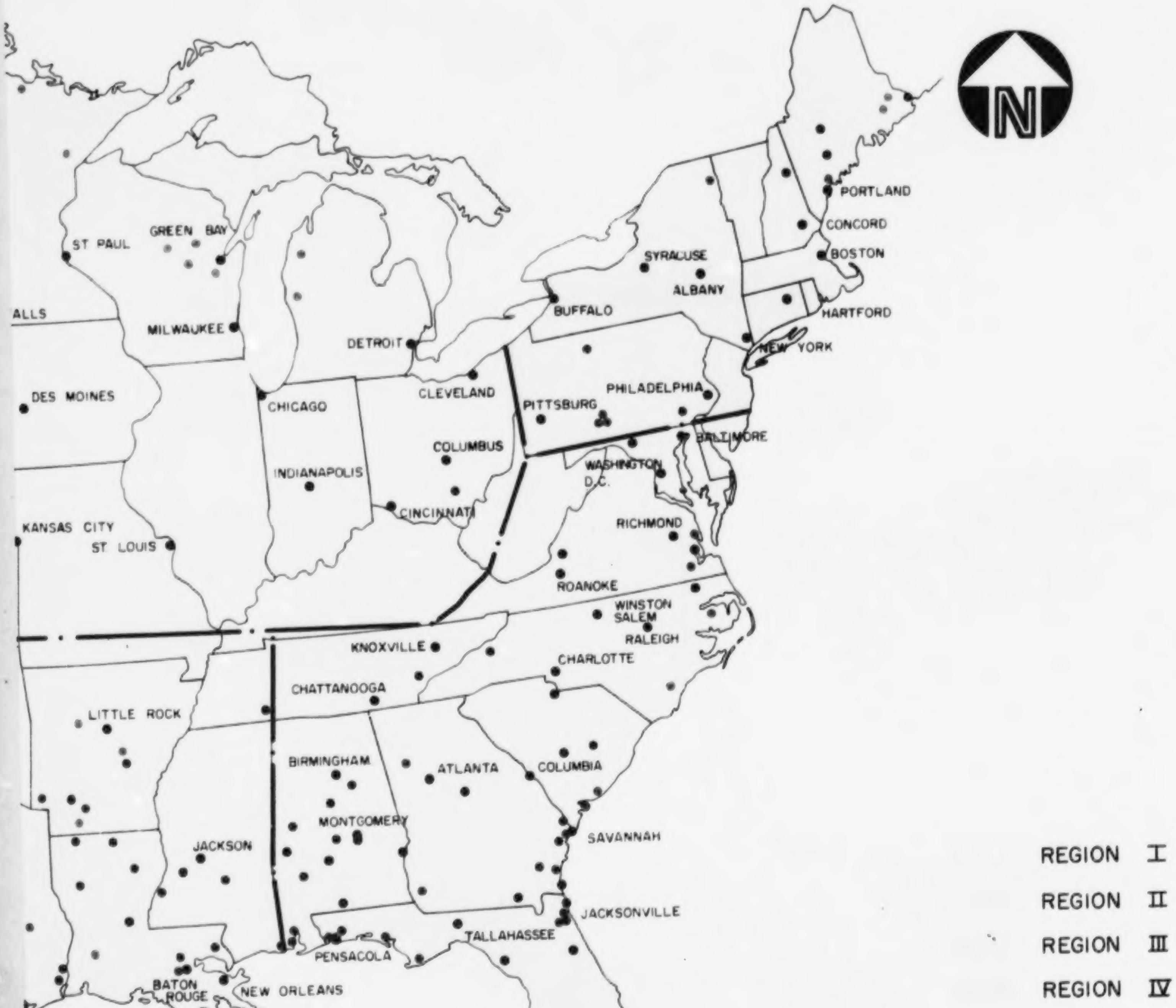
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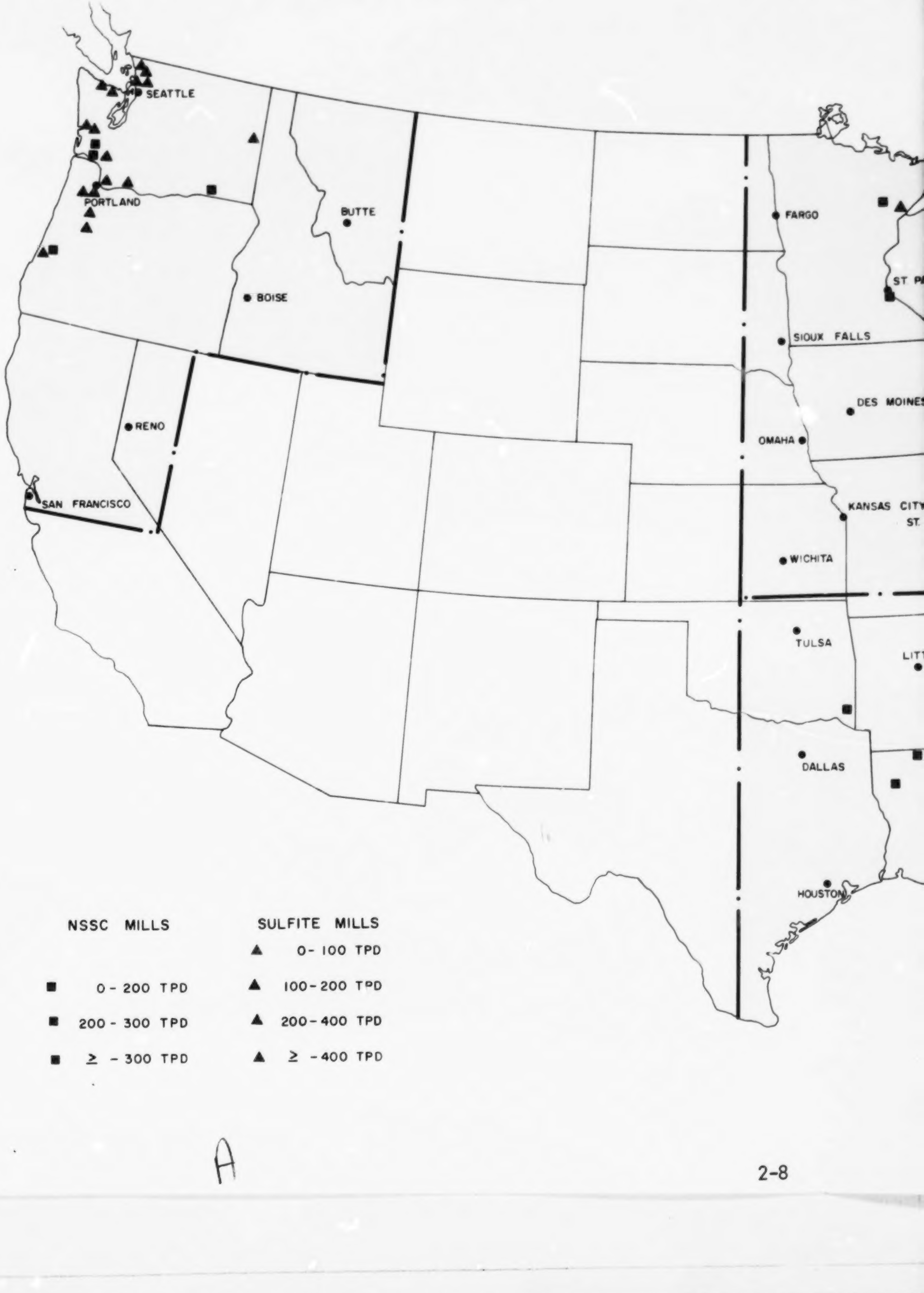
B

REGIONAL DISTRIBUTION OF KRAFT PULP MILLS IN THE U. S.

FIGURE 2-1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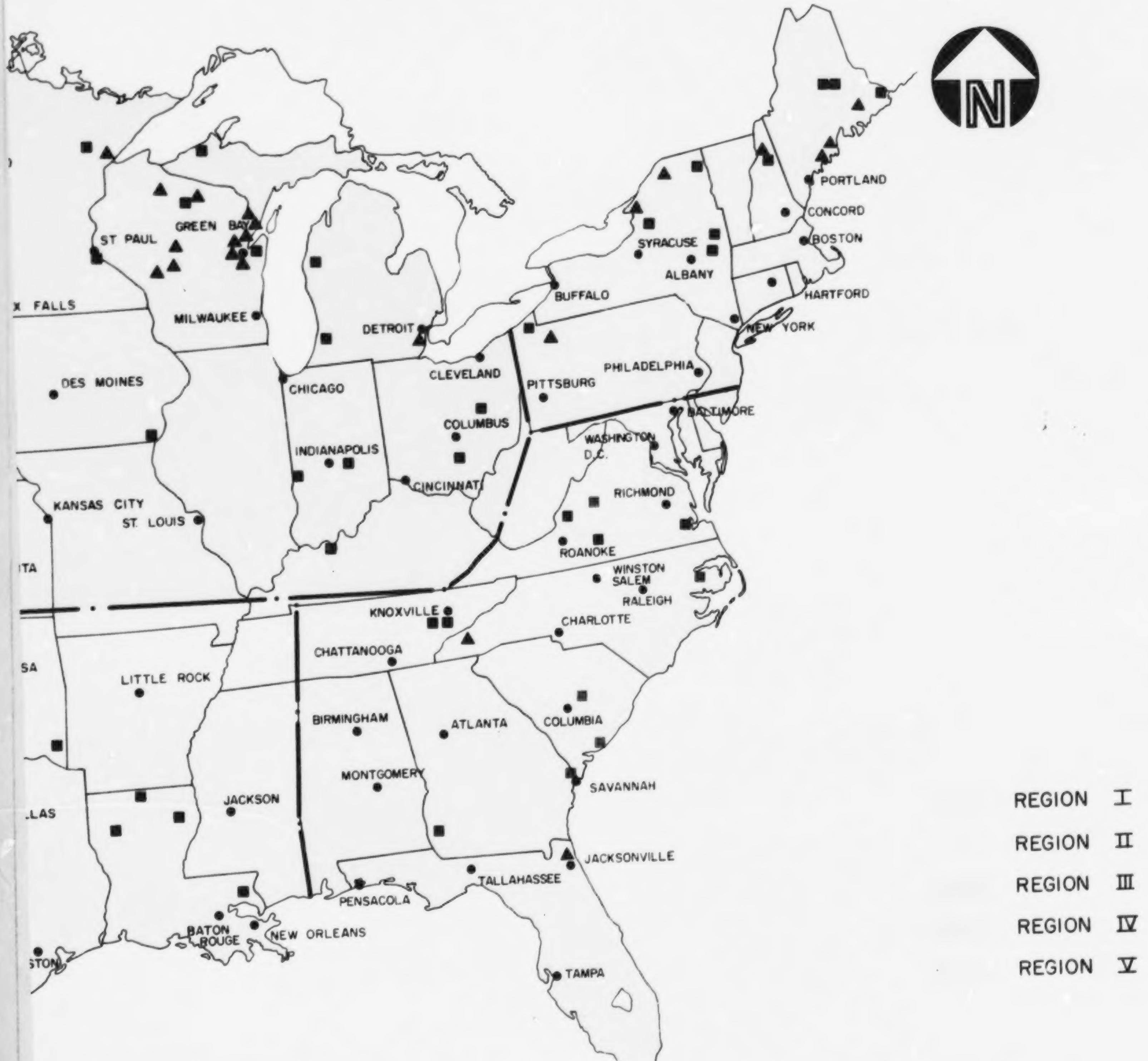
■ 0 - 200 TPD
■ 200 - 300 TPD
■ ≥ - 300 TPD
SULFITE MILLS
▲ 0 - 100 TPD
▲ 100 - 200 TPD
▲ 200 - 400 TPD
▲ ≥ - 400 TPD

A



N
PORTLAND
CONCORD
BOSTON
HARTFORD
NEW YORK
SYRACUSE
ALBANY
BUFFALO
PHILADELPHIA
BALTIMORE
WASHINGTON D.C.
RICHMOND
ROANOKE
WINSTON SALEM
RALEIGH
CHARLOTTE
COLUMBIA
SAVANNAH
JACKSONVILLE
TALLAHASSEE
TAMPA
ATLANTA
CHATTANOOGA
KNOXVILLE
BIRMINGHAM
MONTGOMERY
PENSACOLA
JACKSON
NEW ORLEANS
BATON ROUGE
LITTLE ROCK
ST. LOUIS
KANSAS CITY
DES MOINES
CHICAGO
INDIANAPOLIS
CINCINNATI
COLUMBUS
CLEVELAND
PITTSBURG
DETROIT
MILWAUKEE
GREEN BAY
ST. PAUL
X FALLS
ITA
SA
LLAS
STON
REGION I
REGION II
REGION III
REGION IV
REGION V

REGIONAL DISTRIBUTION OF SULFITE AND NSSC PULP MILLS IN THE U. S.

FIGURE 2-2

β

2.4 FORECASTS

2.4.1 GROWTH AND PROCESS TRENDS

A number of forecasts have been made which attempt to portray the future demand for wood pulp (all grades) in the United States. These forecasts range from a low of 61 million tons per year in 1985 as given by Forest Research Report No. 17, 1965 (3), to a high of 89 million tons per year in 1985 as given by Resources in America, 1963 (4). A middle of the road forecast has been made by the American Paper Institute. Based on these data, plus numerous other sources, and a wealth of in-house knowledge, H. W. Meakin of the J. E. Sirrine Company has projected chemical pulp production through 1985. These projections are reproduced here as Figure 2-3. Historical data are presented in Appendix A (Table A-3).

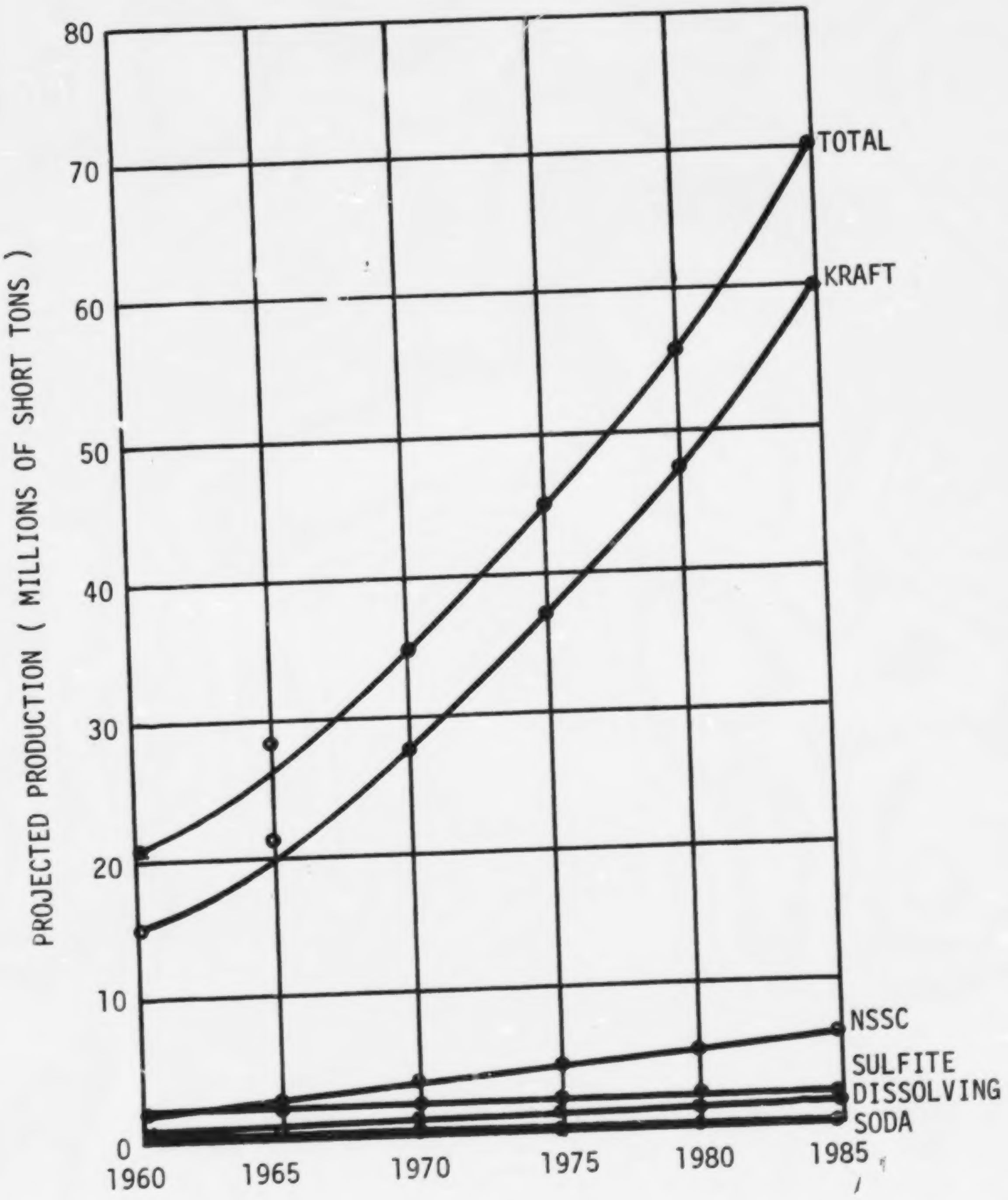
Viewed together, these data show that through 1985, the production of soda pulp and dissolving pulps will remain reasonably constant; sulfite pulp production will decrease slightly; NSSC production will nearly double, and kraft pulp will increase to approximatley 2 1/2 times the 1968 amount.

In 1985, kraft and NSSC processes are expected to dominate chemical pulping in the United States. Kraft production is projected to account for about 85 percent of the chemical (about 70 percent of total wood pulp, all grades, production), and NSSC for about 9 percent of the total chemical pulp production. The total production of chemical pulp is expected to slightly more than double.

Table 2-3 has been included to summarize announced and estimated expansion and phasing-out operations through 1980. Detailed breakdowns of Table 2-3 may be found in Appendix A (Tables A-4, A-5, A-6, and A-7).

FIGURE 2-3

PROJECTION OF PRODUCTION
OF CHEMICAL PULPS IN THE U. S.



YEAR

TABLE 2-3

ANNOUNCED AND ESTIMATED EXPANSION AND PHASING OUT PLANS THROUGH 1980 (a)

I. Current and Planned New Plant Construction as of December 31, 1968[b]

	CAPACITY ADT/DAY		
	Kraft	Sulfite	NSSC
New	5,866 (12)	830 (2)	750 (3)
Expansion	2,135 (5)	0	568 (2)
TOTAL	8,001 (17)	830 (2)	1,318 (5)

II. Estimate of Phased Out Operations

	CAPACITY ADT/DAY			
Time Period	Kraft	Sulfite	NSSC	Soda
In 1968[c]		835 (5)		60 (1)
In 1969-70	205 (1)	503 (3)	235 (1)	
In 1970-80	85 (1)	1,562 (17)		140 (2)
TOTAL	290 (2)	2,900 (25)	235 (1)	200 (3)

(a) Detailed breakdowns of these data may be found in Appendix A-

(b) Figures in ( ) indicate number of mills

(c) These capacity figures for 1968 are not included in Table 2-1.

In addition to the current and planned new plant construction shown above, there are at least twelve proposed or tentative mills in the talking stage of development. These twelve mills would, if brought to production, supply in excess of an additional 3,000 tons per day of pulp.

2.4.2 GEOGRAPHIC DISTRIBUTION TRENDS

Table 2-4 contains information which shows the regional distribution of the industry in 1968 as well as the projected distribution in 1975 and 1980.

It appears to be the concensus of industry representatives on the Pulp Industry Liaison Committee that the distribution of pulp production in the forseeable future (through 1985) will remain essentially as it is today. The projected chemical pulp production shown on Figure 2-3 was, therefore, stratified by region on the basis of this assumption.

There are several factors which influence the decision to locate a pulp mill in a given section of the country. One of these factors is the availability of trees to serve as raw material. Some concern has been expressed by forestry management people (5, 6) that a tightening of the wood supply in the South could occur in the late 1970's. If this were to occur, there could possibly be a shift of production to the West and North. It is felt, however, that by more intensified management of the better forest lands and improved silvaculture, we can grow appreciably more wood and thus satisfy the demands of the wood pulping industry. Thus, it is predicted that the distribution of chemical pulp production by regions will remain substantially as it is today.

TABLE 2-4

PROJECTION OF PULP PRODUCTION BY REGION IN THE U. S.
1968 - 1980

	1968		1975		1980		
REGION	Prod. TPD	% of Indust. Prod.	Prod. TPD	% Growth Over 1968	Prod. TPD	% Growth Over 1968	% Growth Over 1975
Kraft							
Northeast	3,617	5.21	5,582	54.3	6,981	93.0	25.1
Northcentral	2,319	3.34	3,579	54.3	4,476	93.0	25.1
Southeast	36,249	52.21	55,939	54.3	69,962	93.0	25.1
Southcentral	15,850	22.83	24,460	54.3	30,592	93.0	25.1
Northwest	11,393	16.41	17,582	54.3	21,989	93.0	25.1
TOTAL	69,428	100.0	107,142	54.3	134,000	93.0	25.1
Sulfite							
Northeast	1,464	20.49	1,171	(20.0)	1,171	(20.0)	0
Northcentral	1,354	18.96	1,083	(20.0)	1,083	(20.0)	0
Southeast	471	6.59	377	(20.0)	377	(20.0)	0
Southcentral	0	---	0	---	0	---	---
Northwest	3,854	53.96	3,083	(20.0)	3,083	(20.0)	0
TOTAL	7,143	100.00	5,714	(20.0)	5,714	(20.0)	0
NSSC							
Northeast	1,731	17.31	2,324	34.3	2,720	57.1	17.0
Northcentral	3,330	33.30	4,472	34.3	5,233	57.1	17.0
Southeast	2,926	29.26	3,929	34.3	4,598	57.1	17.0
Southcentral	1,283	12.83	1,723	34.3	2,016	57.1	17.0
Northwest	730	7.30	980	34.3	1,147	57.1	17.0
TOTAL	10,000	100.00	13,428	34.3	15,714	57.1	17.0

TPD stands for Tons Per Day
( ) Represents a decline in production

2.5 REFERENCES

1. Staff, "19th Annual World Review," Pulp and Paper, 43(7), 7- , (June 25, 1969).

2. Staff, "Expansion/Modernization/Acquisition Report," Pulp and Paper, 42(51), 43- , (December 16, 1968).

3. "Timber Trends in the United States," Forest Service, U. S. Department of Agriculture, Washington, 1965.

4. "Resources in America's Future," Landsberg, Fishman and Fisher, New York, 1963.

5. Josephson, H.R. (Director, Division of Forest Economics and Marketing Research, Forest Service, USDA), "Availability of Wood Supplies for the Pulp and Paper Industry," Paper presented at 1968 Annual Meeting of Pulp and Raw Materials Division of American Paper Institute, New York, February 20, 1968.

6. Slatin, Benjamin (Economist,API), "Future Demands for Pulp and Paper as They Influence Pulp Wood Requirements," Paper presented at fall meeting of the Southeastern Technical Division of the American Pulp Wood Association, Atlanta, November 21, 1968.

CHAPTER 3

PRESENT PULPING PRACTICES

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CHAPTER 3

PRESENT CHEMICAL PULPING PRACTICES

SUMMARY

In wood pulping, cooking chemicals have the function of dissolving the lignin that bonds the cellulose fiber together. Thus various chemical processes have been developed, using acid, alkaline, or neutral solutions, which delignify with as little destruction of the cellulose as possible. Most of the chemical pulping processes in use today utilize sulfur in some form in the cooking liquor. In bringing about the solution of wood components, the sulfur combines with constituents of the wood to produce gaseous and particulate compounds which may degrade the quality of air or water if discharged into the environment.

Three of the chemical processes (kraft, sulfite, and neutral sulfite semichemical) account for approximately 80 percent of pulp production in the U. S. The choice of the pulping process is determined by the product being made, by the type of wood species available, and by economic considerations. Therefore there is not a free choice in the process to be used. The three processes have been identified as possible sources of particulate and gaseous emissions into the atmosphere. For this reason they are the subject of this study.

To specifically illustrate the air pollution potentials of the industry, flow diagrams representing pulping processes typical of mills producing the majority of the nation's total pulp output were prepared. These simulated flow diagrams include variations of the basic kraft, sulfite, and NSSC processes and are prepared to feature material balances and processes or equipment which will affect the selection of air quality control measures. The power plant energy balances associated with each flow diagram were also prepared to stress the air quality aspect.

Ten flow diagrams are presented for variations of the kraft process, four for sulfite, and three for NSSC. Except for Kraft Flow Diagram No. 10, only process variations utilized by a significant number of mills have been considered. With each flow diagram is presented information on the typical age and location of the type of mill, general data about the process arrangement and assumptions made, plus figures on representative emissions from each source. The flow diagrams are hypothetical, for use in later parts of the study, and none can be identified with a specific mill.

3.1 INTRODUCTION

Three of the chemical processes (kraft, sulfite, and neutral sulfite semichemical) account for approximately 80 percent of pulp production in the U. S. The choice of the pulping process is based on a variety of factors including the product being made, the type of wood species available, and economic considerations. Therefore, there is not a free choice in the process to be used. The three processes have been identified as sources of particulate and gaseous emissions into the atmosphere. For this reason they are the subject of this study.

To specifically illustrate the air pollution potentials of the industry, simulated flow diagrams representing pulping processes typical of mills producing the majority of the nations's total pulp output were prepared. These diagrams include modifications of the basic kraft, sulfite, and NSSC processes and are prepared to feature material balances and processes or equipment which will affect the selection of air quality control measures. The power plant energy balances associated with each flow diagram were also prepared to stress the air quality aspect. Thus, not all unit processes are shown on each flow diagram and where the emissions from two unit processes normally are discharged through the same vent, they may be indicated as one.

Diagrams were prepared to show material balances on the basis of one ton of unbleached air dry pulp. Except for Kraft Flow Diagram No. 10, only process arrangements utilized by a significant number of mills have been considered. Diagrams have not been prepared for the groundwood and soda pulping processes because the former is a mechanical process with very little or no particulate or odorous emissions and the latter is considered obsolete. The flow diagrams are hypothetical and none can be identified with a specific mill.

Combination boilers firing bark plus one or more fossil fuels are common. The emissions from bark firing have been shown separately and are estimated on the basis of debarking 100 percent of the wood supply. For mills receiving wood chips the bark flow will be reduced in proportion to the amount of chips received. Therefore, the oil and coal consumption must be increased accordingly. Since the amount of bark burned varies considerably from mill to mill, the emission estimates from bark firing vary accordingly.

Boiler emissions for both coal and oil have been calculated and are indicated on the flow diagrams. Only one set of these emissions is applicable depending upon the particular fuel in use. In NO case should the coal and oil emissions be added together.

Emissions from a power boiler firing natural gas are not shown, primarily due to lack of space. The power boiler emission when firing natural gas is approximately the same as fuel oil in terms of flue gas volume and weight. Of course, natural gas has no SO_2 or particulate emission.

A sulfur content of two percent was selected as being a reasonable value for both coal and oil and a value which can be easily adjusted for any fuel analysis. See Chapter 8 for further comments regarding fuel consumption and composition.

3.1.1 PURPOSE OF THIS CHAPTER

The purpose of this chapter is to show flow diagrams and power plant energy balances in order to identify and quantify significant particulate and malodorous emissions from major sources in the pulping processes. In subsequent chapters, various pieces of emission control equipment will be studied for these sources with a view to analyzing the cost of this equipment versus its effectiveness in reducing emissions. The power plant energy balances were developed in order to determine the emissions resulting from the generation of the steam required for process and electric power generation.

The quantity and concentration of emissions on all flow diagrams are based on "annual averages" and are not intended to be used for establishing emission standards, guides, or criteria.

3.1.2 DEFINITIONS

The following definitions are provided for the benefit of those not intimately familiar with the chemical wood pulping industry. In some instances definitions are given because of special meanings which a term may have in this report.

Particulate - Any material which exists as a solid in a gas stream at duct conditions and is collected in accordance with IGCI sampling procedures.

Trace - A quantitative expression of emissions which is less than 0.01 pounds per ton of air dry unbleached pulp.

Standard Conditions - 29.92 inches of mercury and 70 degrees F.

Sulfidity - An expression of the percentage makeup of chemicals in kraft cooking liquor obtained by the formula

$$\frac{Na_2S}{Na_2S + NaOH} \times 100$$

where the sodium compounds are expressed as Na_2O.

Yield - The percentage of a specific weight of bone dry wood that is converted to bone dry pulp.

Weak Wash - A liquid stream in the kraft process which results from washing of the lime mud. *It is used mainly for dissolving smelt.*

Smelt - The molten chemicals from the kraft recovery furnace consisting mostly of sodium sulfide and sodium carbonate.

Oxidation Efficiency - The percentage of sodium sulfide in the kraft black liquor which is oxidized by air introduced into the liquor. The Na_2S is usually expressed in grams per liter of black liquor.

Roundwood - Logs as delivered to the mill with bark attached and cut to specified lengths (up to 10 feet).

Board - A heavy sheet made with single or multiple plies of pulp formed on a board machine such as a fourdrinier.

Linerboard - A laminated container board usually made of kraft pulp. It consists of a base sheet which is coarse strong pulp and a top liner sheet which is fine pulp. The top liner gives the container board a more finished exposed sulface than the base sheet.

Top Liner - A sheet, usually produced from kraft pulp, which is added as a laminate to the base sheet to produce linerboard. The pulp may in some cases be bleached.

Base Stock - A sheet, usually produced from unbleached kraft pulp, formed into linerboard on a fourdrinier machine.

Corrugating Medium - A sheet, usually made from NSSC pulp, which is corrugated to form a cushioning layer when attached to a single sheet or between two boards. The corrugating sheet is usually 0.009 inches thick and traditionally is referred to as "9 point."

Bark Boiler - A combustion unit used to produce steam for process or electrical energy which is designed to burn mainly bark and wood residues.

Combination Boiler - A combustion unit used to produce steam for process or electrical energy which is designed to burn bark and at least one other fuel.

Power Boiler - A combustion unit used to produce steam for process or electrical energy which is designed to burn oil, coal, or gas.

Recovery Boiler - A combustion unit used to produce steam for pulping and recovery operations, and to recover the spent chemicals from the cooking liquor.

3.1.3 ROLE OF CHEMICALS IN PULPING

In wood pulping, cooking chemicals have the function of dissolving the lignin that bonds the cellulose fibers together. Ideally, the chemical process should dissolve all of the intercellular cementing constituents and extraneous materials without affecting the cellulose fibers. Unfortunately, the ideal is never achieved. Thus, various chemical processes have been developed using acid, alkaline, or neutral solutions, which delignify with as little destruction of the cellulose as possible. All of the pulping processes of interest in this study utilize sulfur in some form in the cooking liquor. In bringing about the solution of wood components, the sulfur reacts with some of the wood components to produce compounds which may degrade the quality of air or water if released into the environment.

During the early development stages of the wood pulp industry, the use of chemicals was insignificant. But as new chemical pulping techniques came into being and the need for pulp products increased, it created an increasing demand for chemicals. Now the chemical requirements are such that it is an important segment of the chemical industry.

Of the three major chemical pulping processes in use today, the sulfite process was the first. Established in 1867, it used a calcium base cooking acid. The raw materials, sulfur and limestone, were inexpensive and plentiful. Pyrite, in some areas where available, was used instead of sulfur for the formation of sulfur dioxide. In place of limestone, some mills preferred the milk of lime system, which can be calcium hydroxide or a slurry of calcium carbonate.

The use of magnesium as a base for sulfite pulping has been known since 1874. But because of the relatively high cost of the chemical make-up, magnesium oxide, it was not commercialized until 1948. It was at this time that a recovery system was developed permitting the reuse of magnesium oxide and sulfur dioxide that is recovered in the combustion of the spent liquor.

In 1948 a number of calcium-base mills were converted to the use of ammonia. The higher cost of the ammonia-base over the calcium-base is offset by a substantial increase in production because of a 25 percent decrease in the time of a pulping cycle.

Kraft pulping came into being in 1879. It was a modification of the caustic soda system, whereby sodium sulfide was added to the caustic soda cooking liquor. The introduction of the modern recovery furnace in the period 1928-1934 brought about a tremendous increase in the use of kraft pulp. Recovery of cooking chemicals from kraft spent liquor is essential for the kraft process to be cost competitive with other processes. The recovery of chemicals is accomplished by spraying concentrated spent liquor into the recovery furnace, where the organic compounds are burned and an inorganic smelt of sodium sulfide and sodium carbonate is formed. To make up for chemicals lost in the operating cycle, salt cake (sodium sulfate) is usually added to the concentrated spent liquor before it is sprayed into the furnace.

The smelt of sodium sulfide and sodium carbonate flows from the furnace and is dissolved in water to form green liquor. This solution is reacted with slaked lime (calcium hydroxide), converting the green liquor to cooking liquor which is a solution of sodium hydroxide and sodium sulfide. The calcium carbonate created by this reaction is settled out, dewatered, and burned in a lime kiln. The resultant calcium oxide is returned for reaction with the green liquor.

NSSC pulp is produced with a neutral sulfite cooking liquor. This liquor is prepared by reacting excess soda ash or caustic soda with sulfur dioxide producing a solution of pH 8 to 11. The chief use of NSSC pulp is in the production of corrugating medium for board grades.

3.1.4 CHARACTERISITICS OF EACH PROCESS

3.1.4.1 Kraft

The Kraft process produces a dark colored fiber. Therefore, the market for the unbleached pulp is usually limited to its use in board, wrapping, and bag papers. If kraft pulp is to be used in the manufacture of fine white papers, its fibers must be treated additionally in a bleach plant.

Cooking chemicals (caustic soda and sodium sulfide) are expensive to manufacture. Thus their recovery is an economic necessity. During the recovery process, steam is produced from the combustion of the organic materials, adding to the economic benefits of the recovery system.

The presence of caustic soda in the cooking liquor permits the pulping of practically all wood species. The other active chemical, sodium sulfide, creates a chemical reaction during cooking that imparts the strength characteristics to kraft fibers, producing fibers that are stronger than those made from NSSC or sulfite processes. Small amounts of sodium sulfide react with lignin and carbohydrates in the wood to form odorous compounds which may cause a reduction of air quality.

3.1.4.2 Neutral Sulfite Semichemical

This process is mainly used for the production of a high yield pulp having a high crush strength, important for making corrugated board. In addition, it utilizes hardwood species that are not readily adaptable to the other processes. Coniferous woods are considered less desirable for the NSSC process because of a higher chemical consumption during cooking, high lignin content for a given yield, and high energy requirements for refining.

NSSC pulp with high yields (75 to 80 percent) is used mainly in the production of corrugating medium and linerboard. Some is used for blending with kraft pulp for carton board, wrapping papers, et cetera. If cooked to a yield in the range of 55 to 62 percent, NSSC pulps can be bleached and then blended with bleached sulfite for high grade papers.

Recovery of chemicals from the spent liquor is not practiced at the majority of mills and, therefore, may create a water pollution problem. Incineration of the liquor may, in turn, create an emission problem because of sulfur dioxide emission from the incinerator and will depend on the degree of sulfur dioxide recovery.

3.1.4.3 Sulfite

The sulfite pulping process dominated the commercial pulping field from about 1890 to 1930, producing easy bleaching pulps from non-resinous woods. The cooking chemical was calcium bisulfite plus free sulfurous acid. The characteristics of the pulp make it suitable for use in many grades of paper but it is especially suitable for tissues and fine papers.

Recovery of cooking chemicals and the heat values in the spent cooking liquors was not widespread until fairly recent years when the older calcium base has been replaced in many mills by a soluble base such as sodium, magnesium, or ammonium. Sodium and magnesium bases require recovery for economic reasons. Of the two, magnesium is more widespread in its use because on combustion the inorganic constituents break down directly to magnesium oxide and sulfur dioxide which can readily be recycled. Spent liquors from ammonia base pulping may be incinerated with recovery of most of the SO_2. The ammonia burns completely to nitrogen and water vapor.

The trend toward soluble bases has improved the versatility of the sulfite process in terms of wood species which can be pulped; modifications of the process have led to higher yields of pulp from wood, and recovery processes have not only reduced water pollution, but stimulated the development of chemical by-product opportunities unique to this process.

3.1.5 FLOW DIAGRAMS AND BASIS FOR SELECTION

To describe major pulping operations, 17 basic flow diagrams, each with a power plant energy balance, have been prepared. These flow diagrams illustrate the majority of pulping process variations now in use in this country. These are as follows:

3.1.5.1 Kraft (10 Flow Diagrams)

Table 3-1 lists the significant factors which influence the emission of air pollutants from the kraft pulping process and the particular factors that are to be included in each of the ten flow diagrams.

Gas flow quantities from the turpentine condenser and the multiple effect evaporators have been assumed at 35 cubic feet per air dry ton of pulp on all kraft flow diagrams. These may vary between 20 cubic feet to 60 cubic feet and will depend on individual operations of the relief lines.

For purposes of gas volume calculations, the slaker vent has been assumed to be 40 feet high with a diameter of 24 inches for a pulp production of 600 tons per day, giving a flow of 7,000 cubic feet per air dry ton of pulp. This may vary to a low of 4,000 cubic feet per air dry ton. Particulate emissions are unknown, and since most of the particulates drop within the mill area, this source can be considered as a mill nuisance.

3.1.5.2 Neutral Sulfite Semichemical (3 Flow Diagrams)

1. Combination of neutral sulfite semichemical and kraft pulping (continuous digester).
2. Neutral sulfite semichemical without spent liquor recovery (batch digesters).
3. Fluidized bed recovery process (continuous digester).

TABLE 3-1

KRAFT FLOW DIAGRAM FACTORS

PULPING OPERATION	FLOW DIAGRAM NO.									
	1	2	3	4	5	6	7	8	9	10
Batch Digester	X		X		X		X			
Continuous Digester		X		X		X		X	X	X
Concentrated Black Liquor Oxidation				X						
Weak Black Liquor Oxidation		X	X							
No Oxidation	X				X	X	X	X	X	X
Direct Contact Evap. Recovery	X	X	X	X	X	X		X	X	
Venturi Recovery							X			
Bleach Plant		X				X	X	X	X	
No Bleaching	X		X	X	X					X
Lime Kiln Moderate Collection Efficiency	X				X	X	X			
Lime Kiln High Collection Efficiency			X	X				X	X	X
Fluidized Bed Calcining High Collection Efficiency		X								
High Solids Evaporator (63% Solids)										X
Long Tube Vertical Evaporators	X	X	X	X	X	X	X	X	X	
Hardwood (H) or Softwood (S)	S	S	S	S	S	S	S	H	S	S
Percent Yield	53	45	47	47	47	45	45	46	45	47
Product	Base Linerboard	Bleachable Grades	Paper	Top Liner or Paper	Top Liner	Bleachable Grades	Bleachable Grades	Bleachable Grades	Bleachable Grades	Paper

3.1.5.3 Sulfite (4 Flow Diagrams)

1. Magnesium acid sulfite, with recovery (batch digesters).
2. Calcium acid sulfite, without recovery (batch digesters).
3. Magnesium bisulfite (magnefite) with recovery (batch digesters).
4. Ammonium acid sulfite, with liquor incineration (batch digesters).

The flow diagrams have been prepared from an air quality point of view, and are broken down only to the extent necessary to clearly locate and identify emissions.

3.1.6 POWER PLANT ENERGY BALANCE

Power plant energy balances were developed in order to determine the emissions resulting from the generation of the steam required for process and electric power generation. In order to have a common base for comparison, the balances are based on supplying all of the specific steam and electrical requirements for each process with no outside purchase of either electricity or steam. Steam and electrical usage and steam pressures have been assumed which are representative of actual mills utilizing the various processes.

Since pulp mills must either produce dry pulp for shipment or are part of an integrated pulp and paper operation, the energy balances were developed to include the steam and electric loads associated with either a paper machine or pulp dryer. In order to give an indication of the steam and electric requirements of the pulp mill only, a second set of numbers has been included on each heat balance. These pulp mill requirements were arrived at by performing a balance without the electric and steam requirements of a paper machine or pulp dryer.

The amount of steam generated by the recovery boilers is calculated based on burning the black liquor solids at an efficiency of 60 percent. The steam generated from bark burning is based on an efficiency of 70 percent. The remaining steam is supplied by the power boiler(s) at an efficiency of 85 percent for coal or fuel oil. The quantity of coal or fuel oil required for the power boiler(s) is used as a basis to calculate the air emissions shown on the flow diagrams. In no case should the coal and oil emissions be added together.

3.2 KRAFT

3.2.1 GENERAL DESCRIPTION OF KRAFT PROCESS

The word kraft is derived from a Swedish word which means "strong," because kraft fibers are stronger than those produced by either the NSSC or the sulfite processes.

The kraft process involves the cooking of wood chips in either a batch or continuous digester, under pressure, in the presence of a cooking liquor. The kraft cooking liquor contains sodium hydroxide and sodium sulfide, the hydroxide being the reagent that dissolves the lignin. During the cooking reaction the hydroxide is consumed and the sodium sulfide serves to buffer and sustain the cooking reaction. At the same time, small amounts of sulfide react with lignin in the wood giving rise to the odors characteristic of kraft mills.

Upon the completion of the cook, the residual pressure within the digester is used to discharge the pulp into a blow tank. Gases and flash steam released in the tank are vented through a condenser, where heat is recovered and the condensible vapors removed. The noncondensible gases, which are a source of malodors, are either confined and treated or released to the atmosphere. At the same time the pulp in the blow tank is being diluted and pumped to washers where the spent chemicals and the organics from the wood are separated from the fibers.

The spent chemicals and the organics, called black liquor, are then concentrated in multiple-effect evaporators and/or direct-contact evaporators for subsequent burning. A solids content of 60-70 percent in the black liquor is a requirement for combustion in the recovery furnace.

During evaporation of the black liquor in the multiple-effect evaporators volatile malodorous gases are released. These gases escape where entrained gases and vapors are drawn off by a vacuum system. In order to eliminate the venting of these gases to the atmosphere, they can be confined and destroyed.

The black liquor may be concentrated further in a direct-contact evaporator using hot flue gases from the recovery furnace. These hot gases, containing carbon dioxide, react with sulfur compounds in the black liquor leading to the release of hydrogen sulfide. Prior oxidation of the black liquor will reduce the sulfide content of the liquor and, hence, the amount of hydrogen sulfide released.

The concentrated black liquor is then sprayed into the recovery furnace, where the organic content supports combustion. The inorganic compounds, consisting of the cooking chemicals, fall to the bottom of the furnace where chemical reactions occur in a reducing atmosphere. The chemicals are withdrawn from the furnace as a molten smelt, which is dissolved in a smelt dissolving tank to form a solution called "green liquor." The green liquor is then pumped from the smelt dissolving tank, treated with slaked lime, and then clarified. The resulting liquor, referred to as "white liquor," is the cooking liquor used in the digesters.

There are chemical losses from the kraft process, through air emissions, mill water effluent, and with the finished product. These losses must be made up with purchased chemicals, usually salt cake (sodium sulfate). The name "sulfate" process is derived from this make-up chemical.

3.2.2 BASIC DESCRIPTION OF KRAFT FLOW DIAGRAMS

The basic assumptions which were made in the development of each kraft flow diagram, the diagram itself, and the energy balance are presented on the following pages.

KRAFT DIAGRAM NO. 1

TYPICAL LOCATION

Southern United States.

TYPICAL AGE OF EQUIPMENT

Over 15 years

GENERAL

This flow diagram is based on production of base stock and should be used with Flow Diagram No. 5, top liner stock, to illustrate air emissions from a linerboard pulp mill. Flow Diagram No. 1 depicts a relatively old mill, using batch digesters. A comparison of emissions from the lime kiln is shown when using clean water make-up in the causticizing plant versus the use of multiple-effect evaporator combined condensate.

There are perhaps 30 to 40 mills in the United States that are illustrated by Flow Diagram No. 1. These must be combined with those mills represented by Flow Diagram No. 5.

EMISSIONS

The following emissions have been selected as representative, based on the range of emissions presented in Chapter 4.

POUNDS PER AIR DRY TON OF UNBLEACHED PULP

Location	H_2S	RSH, RSR, RSSR	SO_2	Particulate
Blow Tank Accumulator	0.10	2.90	0	0
Washers and Screens	0.01	0.12	0	0
M. E. Evaporators	0.50	0.4	0	0
R. Boiler and D. C. Evaporators	10.7	1.4	3.4	81.8
After Precipitator	10.7	1.4	3.4	9.8
Smelt Dissolving Tank	0.02	0.05	0	4.0
Slaker	0	0	0	Unknown

KRAFT DIAGRAM NO. 1
(Continued)

EMISSIONS - Continued

POUNDS PER AIR DRY TON OF UNBLEACHED PULP

Location	H_2S	RSH, RSR, RSSR	SO_2	Particulate
Lime Kiln	1.2	1.25	0.1	23.5
Lime Kiln Scrubber	0.48	0.50	Trace	4.3
Turpentine Condenser	0.01	0.51	0	0

ASSUMPTIONS

The following assumptions have been made in developing the flow diagram:

A. PULP MILL

1. Pulp Yield = 53 Percent.
2. Cooking liquor charge per air dry ton of brown stock equals 5,050 pounds of which 780 pounds are chemical solids. Sulfidity equals 27 percent.
3. Black liquor solids from first stage washer equals 15 percent and from M. E. Evaporators equals 50 percent solids.

B. RECOVERY

1. Unit operating at 15 percent excess air at economizer outlet.
2. Particulate matter of 8 grains per SDCF leaving the economizer.
3. Efficiency of 50 percent on particulate removal in the direct contact evaporators.
4. Steam shatter jets utilized on smelt spouts.
5. Design efficiency of 95 percent for precipitator with an annual average operating efficiency of 88 percent.

KRAFT DIAGRAM NO. 1
(Continued)

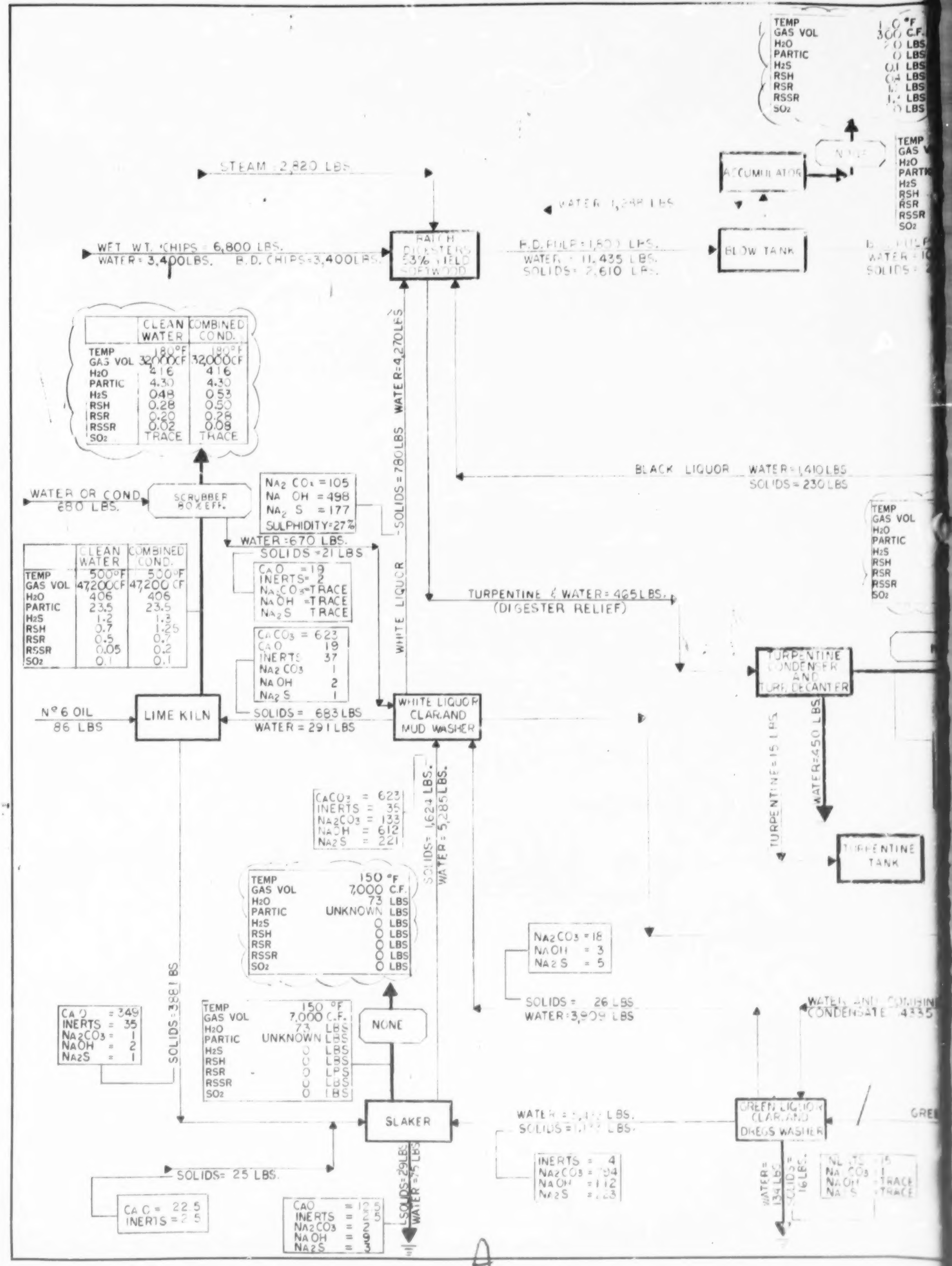
ASSUMPTIONS - Continued

C. CAUSTICIZING

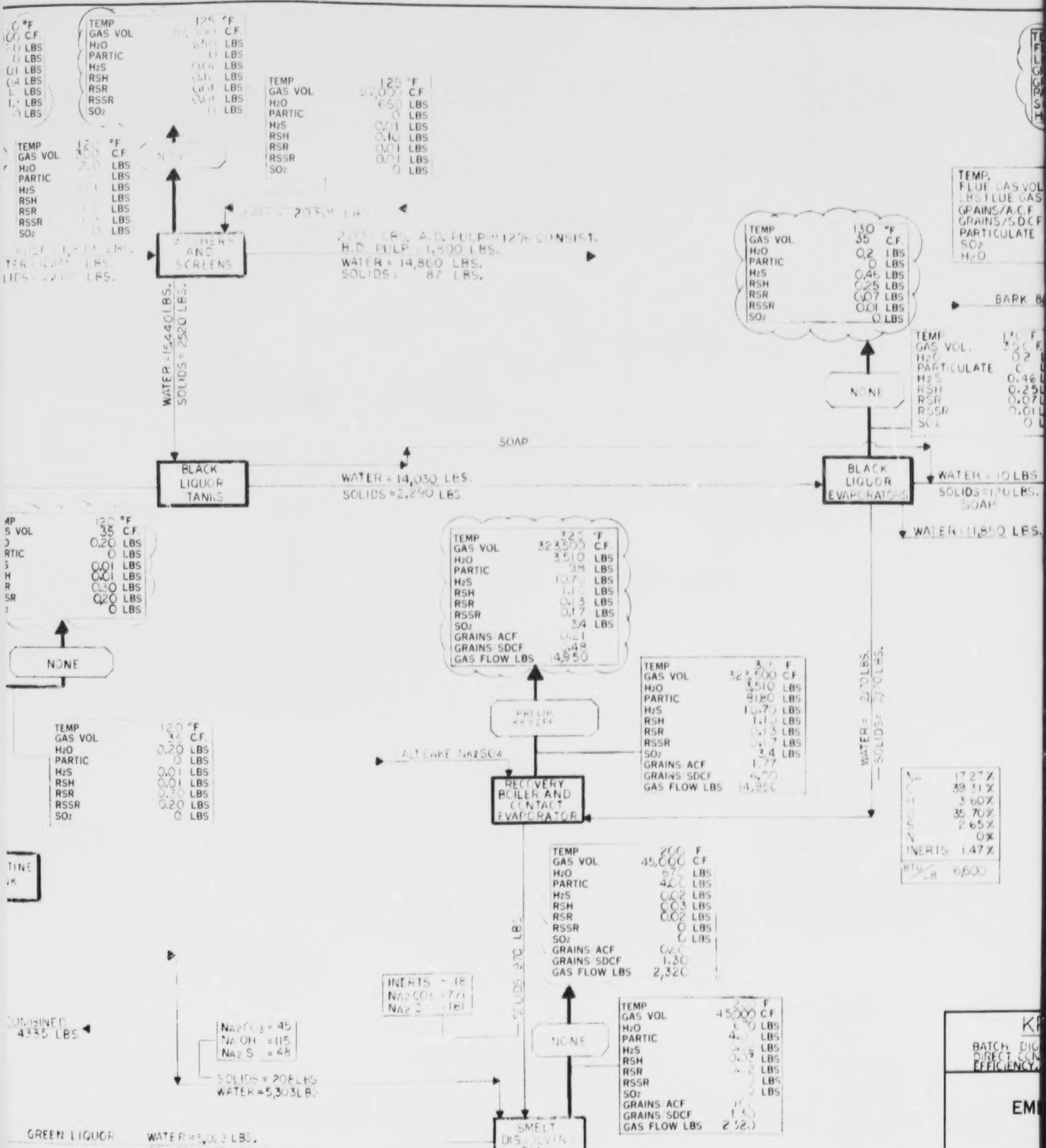
1. Lime kiln SO_2 emission has been assumed at 0.1 pound per air dry ton of pulp.
2. Lime kiln scrubber efficiency assumed to be 80 percent on lime and 60 percent efficiency on soda.

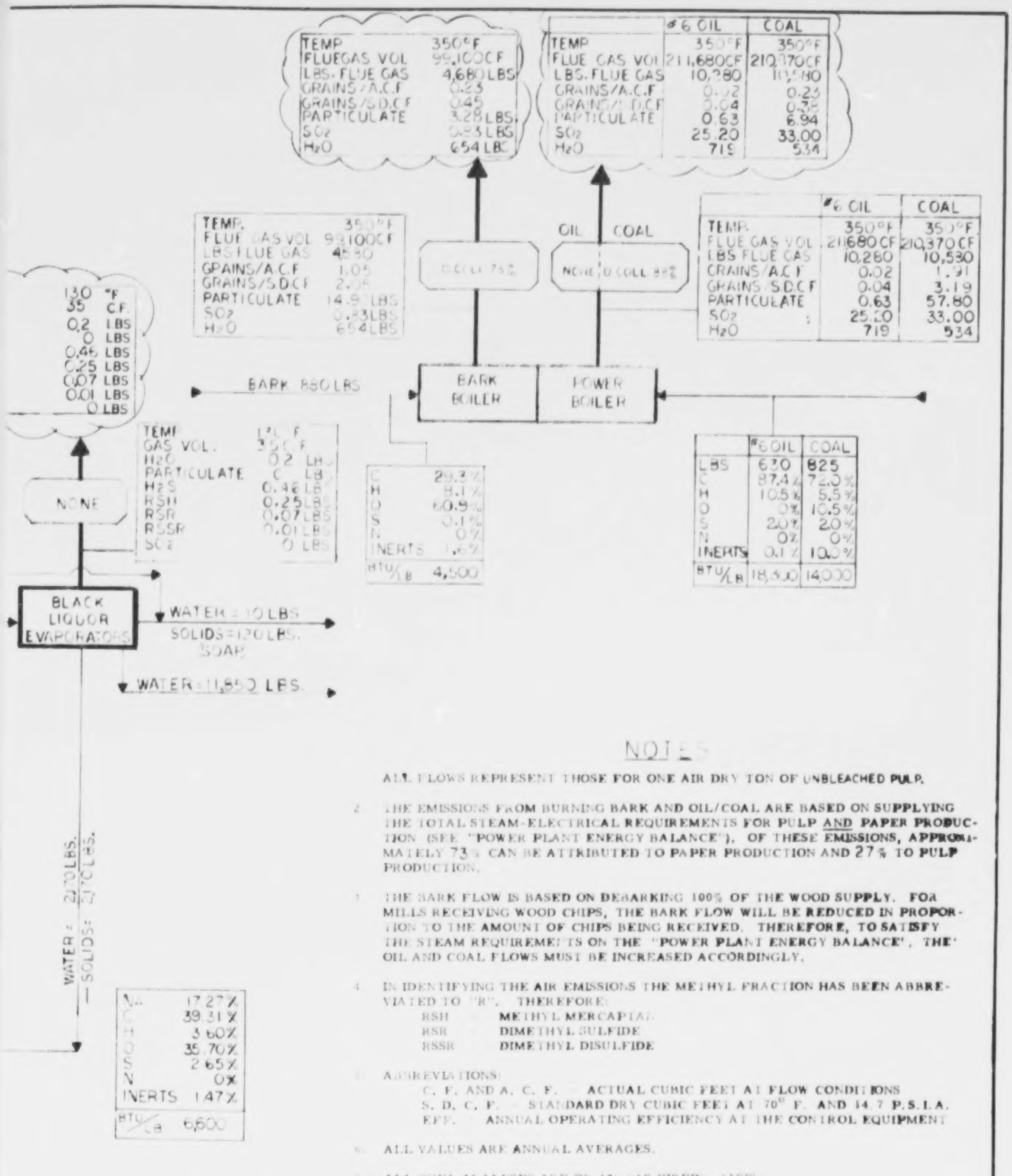
D. POWER PLANT

1. Excess air: 30 percent for bark, 10 percent for oil and 20 percent for coal.
2. Bark burning equipment is a low set spreader stoker.
3. Reinjection from dust collector: 50 percent for bark and 0 percent for coal.
4. Design efficiencies for dust collector: 80 percent on bark and 90 percent on coal with annual operating efficiencies of 78 percent and 88 percent, respectively.
5. Bark is burned 24 hours per day at a controlled rate.
6. Unburned combustible in refuse leaving the dust collector is 40 percent.
7. Coal firing based on pulverized coal or spreader stoker.


STEAM = 2,820 LBS
WET WT. CHIPS = 6,800 LBS.
WATER = 3,400 LBS.
B.D. CHIPS = 3,400 LBS.
BATCH DIGESTERS 53% YIELD SOFTWOOD
B.D. PULP = 1,800 LBS.
WATER = 11,435 LBS.
SOLIDS = 2,610 LBS.
WATER 1,288 LBS
BLOW TANK
ACCUMULATOR
NONE
TEMP
GAS VOL
H2O
PARTIC
H2S
RSH
RSR
RSSR
SO2
°F
300 C.F.
0 LBS
0 LBS
0.1 LBS
0.4 LBS
LBS
LBS
0 LBS
B.D. PULP
WATER
SOLIDS
CLEAN WATER
COMBINED COND.
TEMP 180°F 180°F
GAS VOL 32,000CF 32,000CF
H2O 416 416
PARTIC 4.30 4.30
H2S 0.48 0.53
RSH 0.28 0.50
RSR 0.20 0.28
RSSR 0.02 0.08
SO2 TRACE TRACE
WATER OR COND. 680 LBS.
SCRUBBER 80% EFF.
NA2 CO3 = 105
NA OH = 498
NA2 S = 177
SULPHIDITY = 27%
WATER = 670 LBS.
SOLIDS = 21 LBS
CAO = 19
INERTS = 2
NA2CO3 = TRACE
NAOH = TRACE
NA2S TRACE
TEMP 500°F 500°F
GAS VOL 47,200CF 47,200 CF
H2O 406 406
PARTIC 23.5 23.5
H2S 1.2 1.3
RSH 0.7 1.25
RSR 0.5 0.7
RSSR 0.05 0.2
SO2 0.1 0.1
CACO3 = 623
CAO 19
INERTS 37
NA2 CO3 1
NA OH 2
NA2 S 1
SOLIDS = 683 LBS
WATER = 291 LBS
Nº 6 OIL 86 LBS
LIME KILN
WHITE LIQUOR
SOLIDS = 780 LBS WATER = 4,270 LBS
WHITE LIQUOR CLAR. AND MUD WASHER
BLACK LIQUOR WATER = 1,410 LBS
SOLIDS = 230 LBS
TURPENTINE & WATER = 465 LBS.
(DIGESTER RELIEF)
TURPENTINE CONDENSER AND TURP. DECANTER
TURPENTINE = 15 LBS.
WATER = 450 LBS.
TURPENTINE TANK
CACO3 = 623
INERTS = 35
NA2CO3 = 133
NAOH = 612
NA2S = 221
SOLIDS = 1,624 LBS.
WATER = 5,285 LBS.
TEMP 150 °F
GAS VOL 7,000 C.F.
H2O 73 LBS
PARTIC UNKNOWN LBS
H2S 0 LBS
RSH 0 LBS
RSR 0 LBS
RSSR 0 LBS
SO2 0 LBS
NONE
TEMP 150 °F
GAS VOL 7,000 C.F.
H2O 73 LBS
PARTIC UNKNOWN LBS
H2S 0 LBS
RSH 0 LBS
RSR 0 LPS
RSSR 0 LBS
SO2 0 LBS
SOLIDS = 388 LBS
CAO = 349
INERTS = 35
NA2CO3 = 1
NAOH = 2
NA2S = 1
SLAKER
SOLIDS = 25 LBS.
CAO = 22.5
INERTS = 2.5
SOLIDS = 29 LBS
WATER = 75 LBS
CAO = 12.5
INERTS = 2.5
NA2CO3 = 2
NAOH = 9
NA2S = 3
NA2CO3 = 18
NAOH = 3
NA2S = 5
SOLIDS = 26 LBS
WATER = 3,909 LBS
WATER = LBS.
SOLIDS = LBS.
INERTS = 4
NA2CO3 = 794
NAOH = 112
NA2S = 223
GREEN LIQUOR CLAR. AND DREGS WASHER
WATER AND COMBINED CONDENSATE 4335
WATER = 134 LBS
SOLIDS = 16 LBS.
INERTS = 15
NA2CO3 = 1
NAOH = TRACE
NA2S = TRACE
GREEN

A


KNOTTERS AND SCREENS
2,031 LBS. A.D. PULP @ 12% CONSIST.
B.D. PULP = 1,800 LBS.
WATER = 14,860 LBS.
SOLIDS = 87 LBS.
WATER = 15,440 LBS.
SOLIDS = 2520 LBS.
BLACK LIQUOR TANKS
SOAP
WATER = 14,030 LBS.
SOLIDS = 2,290 LBS.
BLACK LIQUOR EVAPORATORS
WATER = 10 LBS
SOLIDS = 170 LBS.
SOAP
WATER = 11,850 LBS.
WATER = 2,170 LBS.
SOLIDS = 2,170 LBS.
BARK
NONE
PRECIP
RECOVERY BOILER AND CONTACT EVAPORATOR
SALT CAKE Na2SO4
SMELT DISSOLVING TANK
SOLIDS = 208 LBS
WATER = 5,303 LBS.
Na2CO3 = 45
NaOH = 115
Na2S = 48
INERTS = 18
Na2CO3 = 770
Na2S = 161
COMBINED 4335 LBS.
GREEN LIQUOR
WATER = 5,023 LBS.
SOLIDS = 1,175 LBS.
INERTS = 19
Na2CO3 = 813
NaOH = 115
Na2S = 228
TEMP
GAS VOL
H2O
PARTIC
H2S
RSH
RSR
RSSR
SO2
GRAINS/ACF
GRAINS/SDCF
GAS FLOW LBS
325 °F
323,500 C.F.
3,510 LBS
10.70 LBS
14,950
200 °F
45,000 C.F.
2,320
130 °F
35 C.F.
TEMP.
FLUE GAS VOL
LBS FLUE GAS
GRAINS/A.C.F
GRAINS/S.D.C.F
PARTICULATE
SO2
H2O
Na 17.27%
C 39.31%
H 3.60%
O 35.70%
S 2.65%
N 0%
INERTS 1.47%
BTU/LB 6,600
LEGEND
PROCESS EQUIPMENT
CONTROL EQUIPMENT
AIR EMISSIONS
BATCH DIG
DIRECT CON
EFFICIENCY
EM
DE
CO
ENVIRON
B


TEMP 350°F
FLUEGAS VOL 99,100CF
LBS. FLUE GAS 4,680 LBS
GRAINS/A.C.F 0.23
GRAINS/S.D.C.F 0.45
PARTICULATE 3.28 LBS.
SO2 0.83 LBS
H2O 654 LBS
#6 OIL COAL
TEMP 350°F 350°F
FLUE GAS VOL 211,680CF 210,370CF
LBS. FLUE GAS 10,280 10,580
GRAINS/A.C.F 0.02 0.23
GRAINS/S.D.C.F 0.04 0.38
PARTICULATE 0.63 6.94
SO2 25.20 33.00
H2O 719 534
TEMP. 350°F
FLUE GAS VOL 99,100CF
LBS FLUE GAS 4580
GRAINS/A.C.F 1.05
GRAINS/S.D.C.F 2.05
PARTICULATE 14.92LBS
SO2 0.83LBS
H2O 654LBS
OIL
COAL
D.COLL 75%
NONE
D.COLL 88%
#6 OIL COAL
TEMP. 350°F 350°F
FLUE GAS VOL 211680 CF 210,370 CF
LBS FLUE GAS 10,280 10,530
GRAINS/A.C.F 0.02 1.91
GRAINS/S.D.C.F 0.04 3.19
PARTICULATE 0.63 57.80
SO2 25.20 33.00
H2O 719 534
130 °F
35 C.F.
0.2 LBS
0 LBS
0.46 LBS
0.25 LBS
0.07 LBS
0.01 LBS
0 LBS
BARK 850 LBS
BARK BOILER
POWER BOILER
TEMP 130°F
GAS VOL. 35 C.F.
H2O 0.2 LBS
PARTICULATE 0 LBS
H2S 0.46 LBS
RSH 0.25 LBS
RSR 0.07 LBS
RSSR 0.01 LBS
SO2 0 LBS
NONE
C 29.3%
H 8.1%
O 60.9%
S 0.1%
N 0%
INERTS 1.6%
BTU/LB 4,500
#6 OIL COAL
LBS 630 825
C 87.4% 72.0%
H 10.5% 5.5%
O 0% 10.5%
S 2.0% 2.0%
N 0% 0%
INERTS 0.1% 10.0%
BTU/LB 18,500 14,000
BLACK LIQUOR EVAPORATORS
WATER = 10 LBS
SOLIDS = 120 LBS.
SOAP
WATER = 11,850 LBS.
WATER = 2170 LBS.
SOLIDS = 2170 LBS.
NA 17.27%
C 39.31%
H 3.60%
O 35.70%
S 2.65%
N 0%
INERTS 1.47%
BTU/LB 6,600

NOTES

1. ALL FLOWS REPRESENT THOSE FOR ONE AIR DRY TON OF UNBLEACHED PULP.

2. THE EMISSIONS FROM BURNING BARK AND OIL/COAL ARE BASED ON SUPPLYING THE TOTAL STEAM-ELECTRICAL REQUIREMENTS FOR PULP AND PAPER PRODUCTION (SEE "POWER PLANT ENERGY BALANCE"). OF THESE EMISSIONS, APPROXIMATELY 73% CAN BE ATTRIBUTED TO PAPER PRODUCTION AND 27% TO PULP PRODUCTION.

3. THE BARK FLOW IS BASED ON DEBARKING 100% OF THE WOOD SUPPLY. FOR MILLS RECEIVING WOOD CHIPS, THE BARK FLOW WILL BE REDUCED IN PROPORTION TO THE AMOUNT OF CHIPS BEING RECEIVED. THEREFORE, TO SATISFY THE STEAM REQUIREMENTS ON THE "POWER PLANT ENERGY BALANCE", THE OIL AND COAL FLOWS MUST BE INCREASED ACCORDINGLY.

4. IN IDENTIFYING THE AIR EMISSIONS THE METHYL FRACTION HAS BEEN ABBREVIATED TO "R". THEREFORE:
   - RSH — METHYL MERCAPTAN
   - RSR — DIMETHYL SULFIDE
   - RSSR — DIMETHYL DISULFIDE

5. ABBREVIATIONS:
   - C. F. AND A. C. F. — ACTUAL CUBIC FEET AT FLOW CONDITIONS
   - S. D. C. F. — STANDARD DRY CUBIC FEET AT 70° F. AND 14.7 P.S.I.A.
   - EFF. — ANNUAL OPERATING EFFICIENCY AT THE CONTROL EQUIPMENT

6. ALL VALUES ARE ANNUAL AVERAGES.

7. ALL FUEL ANALYSES ARE ON AN "AS FIRED" BASIS.

KRAFT FLOW DIAGRAM NO.1

BATCH DIGESTERS, BASE STOCK, NO OXIDATION, NO [illegible] WITH DIRECT CONTACT EVAPORATOR AND LIME KILN MODERATE COLLECTION EFFICIENCY.

EXHIBIT NO.

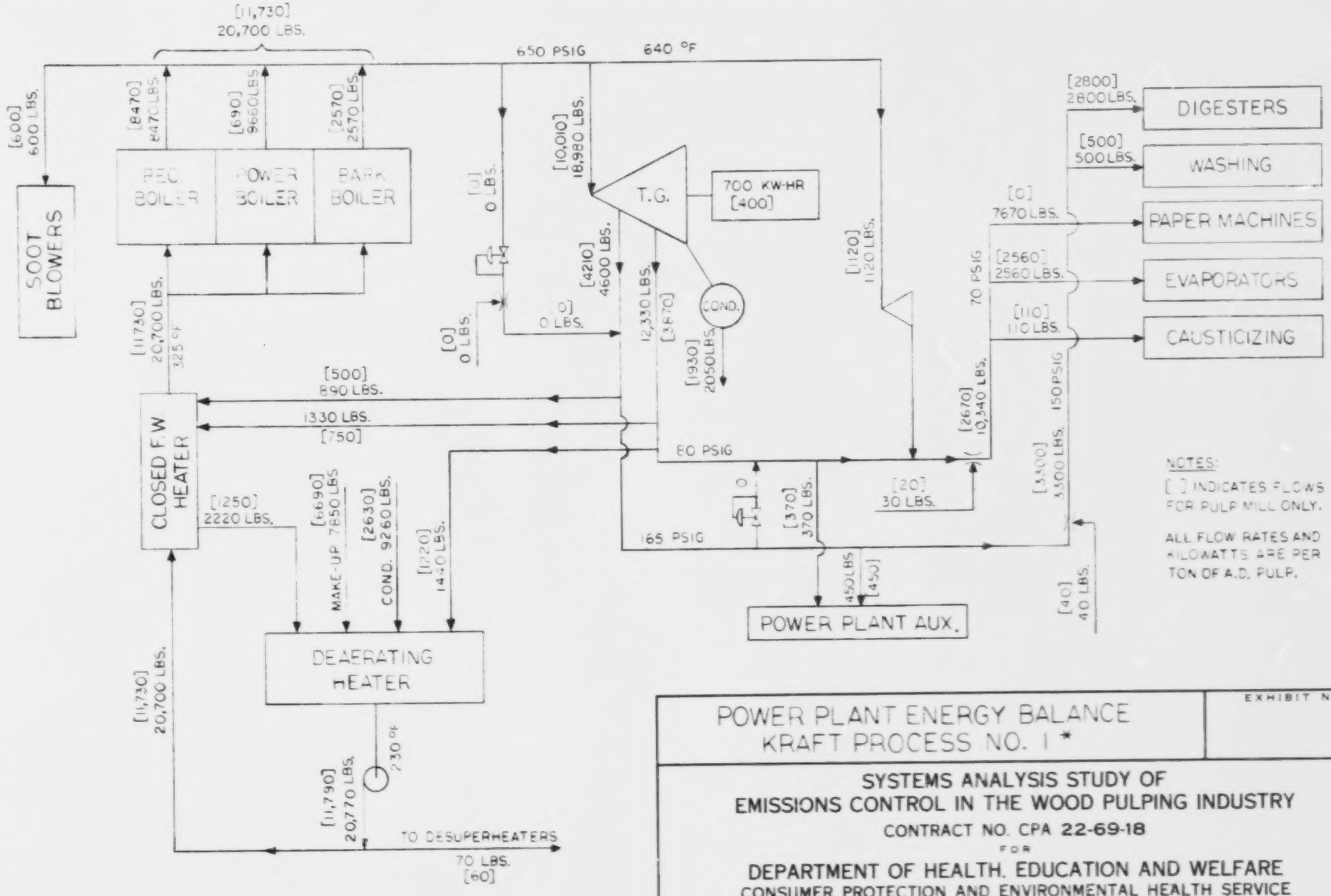
SYSTEMS ANALYSIS STUDY OF EMISSIONS CONTROL IN THE WOOD PULPING INDUSTRY

CONTRACT NO. CPA 22-69-18
FOR
DEPARTMENT OF HEALTH, EDUCATION AND WELFARE
CONSUMER PROTECTION AND ENVIRONMENTAL HEALTH SERVICE
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

EQUIPMENT
EQUIPMENT
SIONS

ENVIRONMENTAL ENGINEERING, INC.
GAINESVILLE, FLORIDA

J. E. SIRRINE COMPANY, ENGINEERS
GREENVILLE, S. C.


[11,730]
20,700 LBS.
650 PSIG
640 °F
[600]
600 LBS.
SOOT BLOWERS
[8470]
8470 LBS.
[690]
9660 LBS.
[2570]
2570 LBS.
REC. BOILER
POWER BOILER
BARK BOILER
[11730]
20,700 LBS.
325 °F
[0]
0 LBS.
[0]
0 LBS.
[0]
0 LBS.
[10,010]
18,980 LBS.
T.G.
700 KW-HR
[400]
[4210]
4600 LBS.
12,330 LBS.
[3870]
COND.
[1930]
2050 LBS.
[1120]
1120 LBS.
[2800]
2800 LBS.
DIGESTERS
[500]
500 LBS.
WASHING
[0]
7670 LBS.
PAPER MACHINES
70 PSIG
[2560]
2560 LBS.
EVAPORATORS
[110]
110 LBS.
CAUSTICIZING
[2670]
10,340 LBS.
150 PSIG
[500]
890 LBS.
1330 LBS.
[750]
80 PSIG
CLOSED F.W. HEATER
[1250]
2220 LBS.
[6690]
MAKE-UP 7850 LBS
[2630]
COND. 9260 LBS.
[1220]
1440 LBS.
0
[370]
370 LBS.
[20]
30 LBS.
[3300]
3300 LBS.
165 PSIG
450 LBS
[450]
POWER PLANT AUX.
[40]
40 LBS.
DEAERATING HEATER
[11,730]
20,700 LBS.
[11,790]
20,770 LBS.
230 °F
TO DESUPERHEATERS
70 LBS.
[60]

NOTES:

[ ] INDICATES FLOWS FOR PULP MILL ONLY.

ALL FLOW RATES AND KILOWATTS ARE PER TON OF A.D. PULP.

* BATCH DIGESTERS, BASE STOCK, NO OXIDATION, NO BLEACHING, DIRECT CONTACT EVAPORATORS, LIME KILN MODERATE COLLECTION EFFICIENCY.

POWER PLANT ENERGY BALANCE
KRAFT PROCESS NO. 1 *

EXHIBIT NO.

SYSTEMS ANALYSIS STUDY OF
EMISSIONS CONTROL IN THE WOOD PULPING INDUSTRY
CONTRACT NO. CPA 22-69-18
FOR
DEPARTMENT OF HEALTH, EDUCATION AND WELFARE
CONSUMER PROTECTION AND ENVIRONMENTAL HEALTH SERVICE
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

ENVIRONMENTAL ENGINEERING, INC.
GAINESVILLE, FLORIDA

J. E. SIRRINE COMPANY, ENGINEERS
GREENVILLE, S. C.

KRAFT DIAGRAM NO. 2

TYPICAL LOCATION

Western United States.

TYPICAL AGE OF EQUIPMENT

Under five years.

GENERAL

Flow Diagram No. 2 contains a fluidized bed calcining system. This flow diagram is based on production of bleached softwood kraft pulp and depicts a new mill designed with emphasis on odor abatement and reduced particulate emission. It includes treatment of the noncondesible gases from digester and multiple-effect evaporators with bleach plant chlorination effluent. Flow Diagram No. 2 represents a new mill designed for a production of 150 to 500 tons per day. There are perhaps two or three mills in the U. S. that are illustrated by this flow diagram.

EMISSIONS

The following emissions have been selected as representative, based on the range of emissions presented in Chapter 4.

POUNDS PER AIR DRY TON OF UNBLEACHED PULP

Location	H_2S	RSH, RSR, RSSR	SO_2	Particulate
Washers and Screens	0.08	0.45	0	0
BLO Tower	0.02	0.30	0	0
M. E. Evaporators	0.02	0.18	0	0
R. Boiler and D. C. Evaporators	1.50	Trace	5.4	116
After Precipitator	1.50	Trace	5.4	3.48
After Scrubber	1.51	0.3*	5.4	1.0
Smelt Dissolving Tank	0.02	0.02	0	1.0

KRAFT DIAGRAM NO. 2
(Continued)

EMISSIONS - Continued

Location	H_2S	RSH, RSR, RSSR	SO_2	Particulate
Slaker	0	0	0	unknown
Fluidized Bed Calciner	unknown	unknown	unknown	72.0
Fluidized Bed Calciner Scrubber	unknown	unknown	unknown	0.7
Turpentine Condenser	0.01	0.43	0	0

*Note: This is a result of introducing the weak liquor oxidation system off-gases into the inlet of the scrubber.

ASSUMPTIONS

The following assumptions have been made in developing the flow diagram:

A. PULP MILL

1. Pulp Yield = 45 Percent.
2. Cooking liquor charge per air dry ton of brown stock equals 6,996 pounds of which 996 pounds are chemical solids. Sulfidity equals 25 percent.
3. It is assumed that there is no production of turpentine and soap.
4. Weak liquor oxidation system ahead of the multiple-effect evaporators. Oxidation efficiency assumed at 99 plus percent.

B. BLEACH PLANT

1. It was assumed as a six stage bleach plant with a chlorine emission to atmosphere of one pound per air dry ton of pulp.

C. RECOVERY

1. Unit operating at 15 percent excess air at economizer outlet.

2. Particulate matter of 8 grains per SDCF leaving the economizer.

3. Efficiency of 50 percent on particulate removal in the direct contact evaporators.

4. Steam shatter jets utilized on smelt spouts.

5. Design efficiency of 98 percent for precipitator with an annual average operating efficiency of 97 percent.

6. Efficiency of 70 percent on particulate removal in the scrubber.

D. CAUSTICIZING

1. Particulate emissions from cyclones assumed at 6 tons per day for 165 ADT/Day unbleached pulp.

2. Scrubber efficiency assumed at 99 percent on lime.

E. POWER PLANT

1. Excess air: 30 percent for bark, 10 percent for oil and 20 percent for coal.

2. Bark burning equipment is a low set spreader stoker.

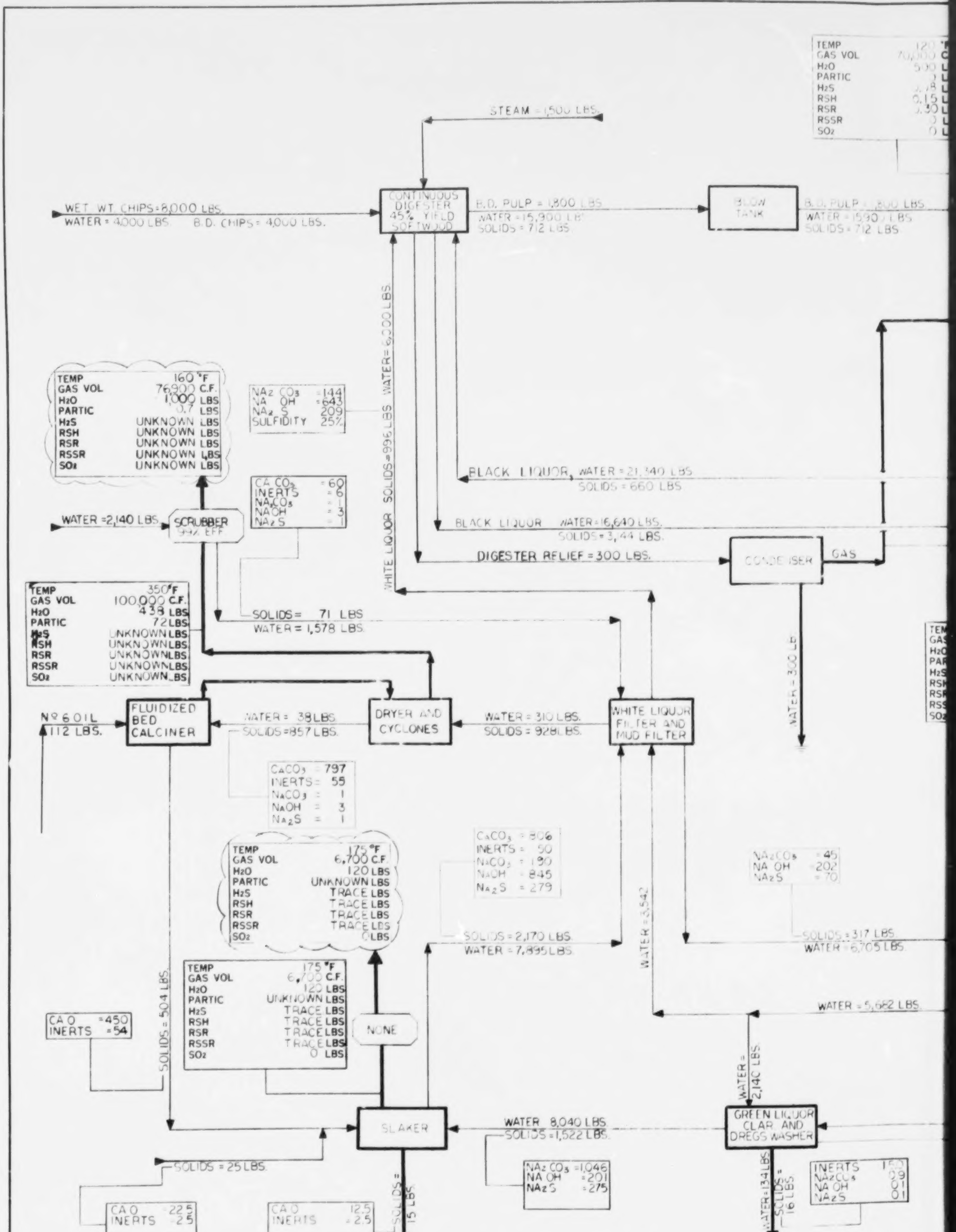
3. Reinjection from dust collector: 50 percent for bark and 0 percent for coal.

4. Design efficiencies for dust collector: 82 percent on bark and 92 percent on coal with annual operating efficiencies of 80 percen and 90 percent, respectively.

5. Bark is burned 24 hours per day at a controlled rate.

6. Unburned combustible in refuse leaving the dust collector is 40 percent.

7. Coal firing based on pulverized coal.


STEAM = 1,500 LBS.
WET WT. CHIPS = 8,000 LBS.
WATER = 4,000 LBS.
B.D. CHIPS = 4,000 LBS.
CONTINUOUS DIGESTER 45% YIELD SOFTWOOD
B.D. PULP = 1,800 LBS.
WATER = 15,900 LBS.
SOLIDS = 712 LBS.
BLOW TANK
B.D. PULP = 1,800 LBS.
WATER = 15900 LBS.
SOLIDS = 712 LBS.
TEMP 120
GAS VOL 70,000
H2O 500
PARTIC 0
H2S 0.78
RSH 0.15
RSR 0.30
RSSR 0
SO2 0
WHITE LIQUOR SOLIDS = 996 LBS. WATER = 6,000 LBS.
NA2 CO3 = 144
NA OH = 643
NA2 S 209
SULFIDITY 25%
TEMP 160 °F
GAS VOL 76,900 C.F.
H2O 1,000 LBS
PARTIC 0.7 LBS
H2S UNKNOWN LBS
RSH UNKNOWN LBS
RSR UNKNOWN LBS
RSSR UNKNOWN LBS
SO2 UNKNOWN LBS
BLACK LIQUOR, WATER = 21,340 LBS
SOLIDS = 660 LBS
CA CO3 = 60
INERTS = 6
NA2CO3 = 1
NAOH = 3
NA2S = 1
WATER = 2,140 LBS.
SCRUBBER 99% EFF
BLACK LIQUOR WATER = 16,640 LBS.
SOLIDS = 3,44 LBS.
DIGESTER RELIEF = 300 LBS.
CONDENSER
GAS
TEMP 350 °F
GAS VOL 100,000 C.F.
H2O 438 LBS
PARTIC 72 LBS
H2S UNKNOWN LBS
RSH UNKNOWN LBS
RSR UNKNOWN LBS
RSSR UNKNOWN LBS
SO2 UNKNOWN LBS
SOLIDS = 71 LBS
WATER = 1,578 LBS.
WATER = 300 LB
Nº 6 OIL 112 LBS.
FLUIDIZED BED CALCINER
WATER = 38 LBS.
SOLIDS = 857 LBS.
DRYER AND CYCLONES
WATER = 310 LBS.
SOLIDS = 928 LBS.
WHITE LIQUOR FILTER AND MUD FILTER
CaCO3 = 797
INERTS = 55
NaCO3 = 1
NaOH = 3
Na2S = 1
TEMP 175 °F
GAS VOL 6,700 C.F.
H2O 120 LBS
PARTIC UNKNOWN LBS
H2S TRACE LBS
RSH TRACE LBS
RSR TRACE LBS
RSSR TRACE LBS
SO2 0 LBS
CaCO3 = 806
INERTS = 50
NaCO3 = 190
NaOH = 845
Na2S = 279
NA2CO3 = 45
NA OH = 202
NA2S = 70
SOLIDS = 2,170 LBS.
WATER = 7,895 LBS.
WATER = 3,542
SOLIDS = 317 LBS.
WATER = 6,705 LBS.
TEMP 175 °F
GAS VOL 6,700 C.F.
H2O 120 LBS
PARTIC UNKNOWN LBS
H2S TRACE LBS
RSH TRACE LBS
RSR TRACE LBS
RSSR TRACE LBS
SO2 0 LBS
NONE
SOLIDS = 504 LBS.
CA O = 450
INERTS = 54
WATER = 5,662 LBS.
WATER = 2,140 LBS.
SLAKER
WATER 8,040 LBS.
SOLIDS = 1,522 LBS.
GREEN LIQUOR CLAR. AND DREGS WASHER
SOLIDS = 25 LBS.
NA2 CO3 = 1,046
NA OH = 201
NA2 S = 275
INERTS 150
NA2CO3 09
NA OH 01
NA2S 01
CA O = 22.5
INERTS = 2.5
CA O 12.5
INERTS = 2.5
SOLIDS = 15 LBS.
WATER = 134 LBS.
SOLIDS = 16 LBS.

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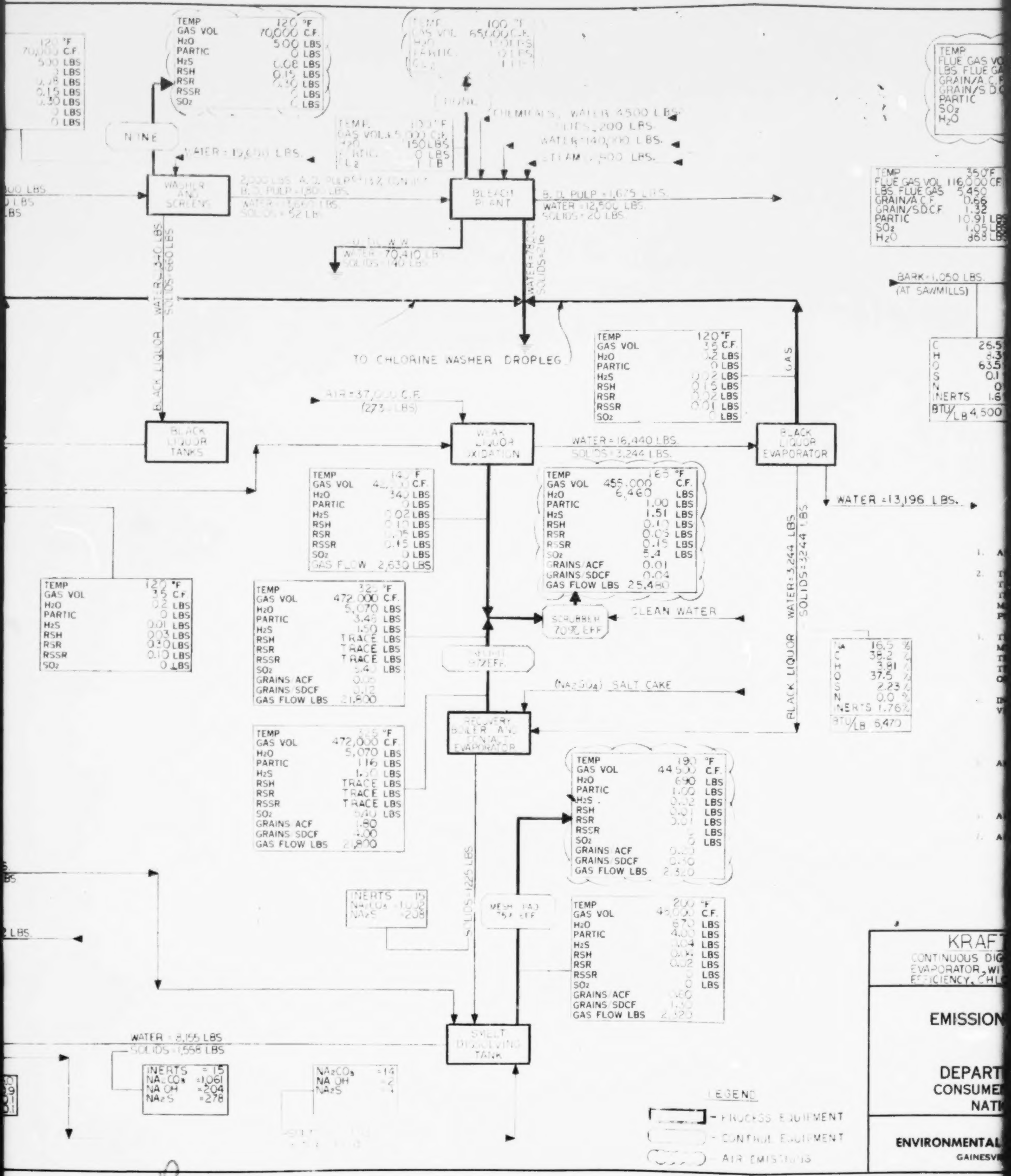

TEMP 120 °F
GAS VOL 70,000 C.F.
H2O 500 LBS
PARTIC 0 LBS
H2S 0.08 LBS
RSH 0.15 LBS
RSR 0.30 LBS
RSSR 0 LBS
SO2 0 LBS
NONE
WATER = 19,650 LBS.
WASHER AND SCREENS
2,000 LBS. A.D. PULP @ 13% CONSIST.
B.D. PULP = 1,800 LBS.
WATER = 13,660 LBS.
SOLIDS = 52 LBS.
WATER = 3,340 LBS
SOLIDS = 660 LBS
BLACK LIQUOR
TEMP 100 °F
GAS VOL 65,000 C.F.
H2O 150 LBS
PARTIC 0 LBS
CL2 1 LB
NONE
TEMP. 100 °F
GAS VOL. 65,000 C.F.
H2O 150 LBS
PARTIC. 0 LBS
CL2 1 LB
CHEMICALS, WATER 4500 LBS.
SOLIDS 200 LBS.
WATER = 140,000 LBS.
STEAM 2,900 LBS.
BLEACH PLANT
B.D. PULP = 1,675 LBS.
WATER = 12,500 LBS.
SOLIDS = 20 LBS.
TO DITCH W.W.
WATER = 70,410 LBS.
SOLIDS = 140 LBS.
WATER = 78,000
SOLIDS = 210
TO CHLORINE WASHER DROPLEG
TEMP 350 °F
FLUE GAS VOL 116,000 CF
LBS FLUE GAS 5,450
GRAIN/A C.F. 0.66
GRAIN/S D.C.F. 1.32
PARTIC 10.91 LBS
SO2 1.05 LBS
H2O 368 LBS
BARK = 1,050 LBS.
(AT SAWMILLS)
C 26.5
H 8.3
O 63.5
S 0.1
N 0
INERTS 1.6
BTU/LB 4,500
TEMP 120 °F
GAS VOL 35 C.F.
H2O 0.2 LBS
PARTIC 0 LBS
H2S 0.02 LBS
RSH 0.15 LBS
RSR 0.02 LBS
RSSR 0.01 LBS
SO2 0 LBS
GAS
AIR = 37,000 C.F.
(2730 LBS)
BLACK LIQUOR TANKS
WEAK LIQUOR OXIDATION
WATER = 16,440 LBS.
SOLIDS = 3,244 LBS.
BLACK LIQUOR EVAPORATOR
WATER = 13,196 LBS.
TEMP 140 °F
GAS VOL 42,000 C.F.
H2O 340 LBS
PARTIC 0 LBS
H2S 0.02 LBS
RSH 0.10 LBS
RSR 0.05 LBS
RSSR 0.15 LBS
SO2 0 LBS
GAS FLOW 2,630 LBS
TEMP 165 °F
GAS VOL 455,000 C.F.
H2O 6,460 LBS
PARTIC 1.00 LBS
H2S 1.51 LBS
RSH 0.10 LBS
RSR 0.05 LBS
RSSR 0.15 LBS
SO2 5.4 LBS
GRAINS/ACF 0.01
GRAINS/SDCF 0.04
GAS FLOW LBS 25,480
TEMP 120 °F
GAS VOL 35 C.F.
H2O 0.2 LBS
PARTIC 0 LBS
H2S 0.01 LBS
RSH 0.03 LBS
RSR 0.30 LBS
RSSR 0.10 LBS
SO2 0 LBS
TEMP 325 °F
GAS VOL 472,000 C.F.
H2O 5,070 LBS
PARTIC 3.48 LBS
H2S 1.50 LBS
RSH TRACE LBS
RSR TRACE LBS
RSSR TRACE LBS
SO2 5.40 LBS
GRAINS/ACF 0.05
GRAINS/SDCF 0.12
GAS FLOW LBS 21,800
SCRUBBER 70% EFF
CLEAN WATER
PRECIP. 97% EFF.
BLACK LIQUOR WATER = 3,244 LBS
SOLIDS = 3244 LBS
NA 16.5 %
C 38.2 %
H 3.81 %
O 37.5 %
S 2.23 %
N 0.0 %
INERTS 1.76 %
BTU/LB 5,470
(NA2SO4) SALT CAKE
RECOVERY BOILER AND CONTACT EVAPORATOR
TEMP 325 °F
GAS VOL 472,000 C.F.
H2O 5,070 LBS
PARTIC 116 LBS
H2S 1.50 LBS
RSH TRACE LBS
RSR TRACE LBS
RSSR TRACE LBS
SO2 5.40 LBS
GRAINS/ACF 1.80
GRAINS/SDCF 4.00
GAS FLOW LBS 21,800
TEMP 190 °F
GAS VOL 44,500 C.F.
H2O 690 LBS
PARTIC 1.00 LBS
H2S 0.02 LBS
RSH 0.01 LBS
RSR 0.01 LBS
RSSR 0 LBS
SO2 0 LBS
GRAINS/ACF 0.20
GRAINS/SDCF 0.40
GAS FLOW LBS 2,320
SOLIDS = 1225 LBS
INERTS 15
NA2CO3 = 1002
NA2S = 208
MESH PAD 75% EFF
TEMP 200 °F
GAS VOL 45,000 C.F.
H2O 670 LBS
PARTIC 4.00 LBS
H2S 0.04 LBS
RSH 0.02 LBS
RSR 0.02 LBS
RSSR 0 LBS
SO2 0 LBS
GRAINS/ACF 0.60
GRAINS/SDCF 1.30
GAS FLOW LBS 2,320
SMELT DISSOLVING TANK
WATER = 8,155 LBS
SOLIDS = 1,558 LBS
INERTS = 15
NA2CO3 = 1061
NA OH = 204
NA2S = 278
NA2CO3 = 14
NA OH = 2
NA2S = 2
2 LBS.
LEGEND
- PROCESS EQUIPMENT
- CONTROL EQUIPMENT
- AIR EMISSIONS
KRAFT
CONTINUOUS DIG
EVAPORATOR, WIT
EFFICIENCY, CHLO
EMISSION
DEPART
CONSUME
NATI
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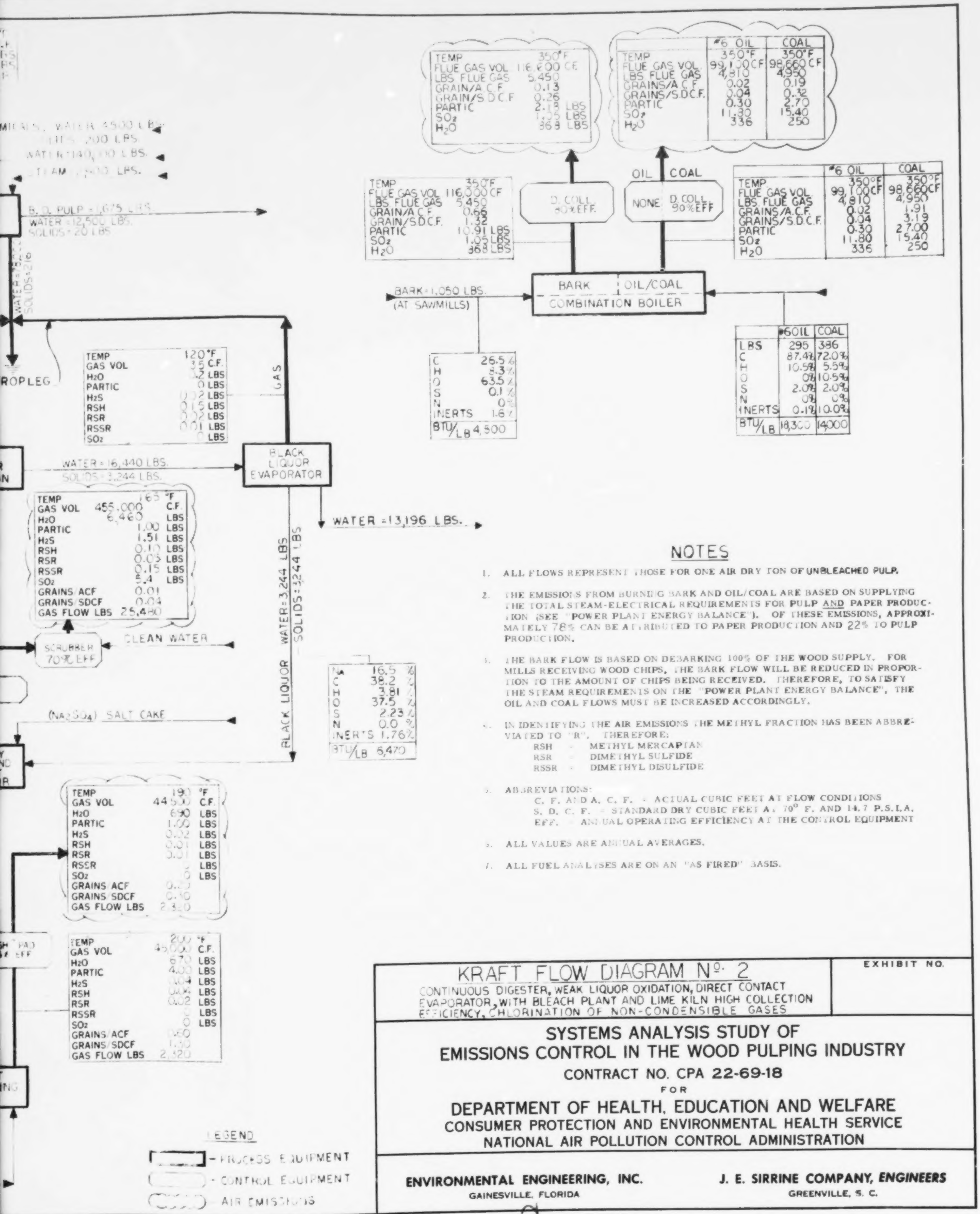
GAINESV

B


TEMP 350°F
FLUE GAS VOL 116,600 CF
LBS FLUE GAS 5,450
GRAIN/A.C.F 0.13
GRAIN/S.D.C.F 0.26
PARTIC 2.18 LBS
SO2 1.05 LBS
H2O 368 LBS
#6 OIL COAL
TEMP 350°F 350°F
FLUE GAS VOL. 99,100CF 98,660CF
LBS FLUE GAS 4,810 4,950
GRAINS/A.C.F 0.02 0.19
GRAINS/S.D.C.F. 0.04 0.32
PARTIC 0.30 2.70
SO2 11.30 15.40
H2O 336 250
OIL COAL
D. COLL. 90%EFF.
NONE
D. COLL. 90%EFF
TEMP 350°F
FLUE GAS VOL 116,000 CF
LBS FLUE GAS 5,450
GRAIN/A.C.F 0.66
GRAIN/S.D.C.F. 1.32
PARTIC 10.91 LBS
SO2 1.05 LBS
H2O 368 LBS
#6 OIL COAL
TEMP 350°F 350°F
FLUE GAS VOL. 99,100CF 98,660CF
LBS FLUE GAS 4,810 4,950
GRAINS/A.C.F. 0.02 1.91
GRAINS/S.D.C.F. 0.04 3.19
PARTIC 0.30 27.00
SO2 11.80 15.40
H2O 336 250
BARK
OIL/COAL
COMBINATION BOILER
BARK=1,050 LBS. (AT SAWMILLS)
C 26.5%
H 8.3%
O 63.5%
S 0.1%
N 0%
INERTS 1.6%
BTU/LB 4,500
#6OIL COAL
LBS 295 386
C 87.4% 72.0%
H 10.5% 5.5%
O 0% 10.5%
S 2.0% 2.0%
N 0% 0%
INERTS 0.1% 10.0%
BTU/LB 18,300 14000
MICALS, WATER 4500 LBS
200 LBS
WATER 140,000 LBS.
B. D. PULP = 1,675 LBS.
WATER = 12,500 LBS.
SOLIDS = 20 LBS
WATER =
SOLIDS = 210
ROPLEG
TEMP 120°F
GAS VOL 35 C.F.
H2O 0.2 LBS
PARTIC 0 LBS
H2S 0.02 LBS
RSH 0.15 LBS
RSR 0.02 LBS
RSSR 0.01 LBS
SO2 0 LBS
GAS
BLACK LIQUOR EVAPORATOR
WATER = 16,440 LBS.
SOLIDS = 3,244 LBS.
WATER = 13,196 LBS.
TEMP 165 °F
GAS VOL 455,000 C.F.
H2O 6,460 LBS
PARTIC 1.00 LBS
H2S 1.51 LBS
RSH 0.10 LBS
RSR 0.05 LBS
RSSR 0.15 LBS
SO2 5.4 LBS
GRAINS/ACF 0.01
GRAINS/SDCF 0.04
GAS FLOW LBS 25,480
SCRUBBER 70% EFF
CLEAN WATER
BLACK LIQUOR WATER = 3,244 LBS
SOLIDS = 3,244 LBS
Na 16.5 %
C 38.2 %
H 3.81 %
O 37.5 %
S 2.23 %
N 0.0 %
INERTS 1.76%
BTU/LB 6,470
(Na2SO4) SALT CAKE
TEMP 190 °F
GAS VOL 44,500 C.F.
H2O 690 LBS
PARTIC 1.00 LBS
H2S 0.02 LBS
RSH 0.01 LBS
RSR 0.01 LBS
RSSR 0 LBS
SO2 0 LBS
GRAINS/ACF 0.20
GRAINS/SDCF 0.40
GAS FLOW LBS 2,320
TEMP 200 °F
GAS VOL 45,000 C.F.
H2O 670 LBS
PARTIC 4.00 LBS
H2S 0.04 LBS
RSH 0.06 LBS
RSR 0.02 LBS
RSSR 0 LBS
SO2 0 LBS
GRAINS/ACF 0.60
GRAINS/SDCF 1.30
GAS FLOW LBS 2,320
PAD EFF
LEGEND
- PROCESS EQUIPMENT
- CONTROL EQUIPMENT
- AIR EMISSIONS

NOTES

1. ALL FLOWS REPRESENT THOSE FOR ONE AIR DRY TON OF UNBLEACHED PULP.
2. THE EMISSIONS FROM BURNING BARK AND OIL/COAL ARE BASED ON SUPPLYING THE TOTAL STEAM-ELECTRICAL REQUIREMENTS FOR PULP AND PAPER PRODUCTION (SEE "POWER PLANT ENERGY BALANCE"). OF THESE EMISSIONS, APPROXIMATELY 78% CAN BE ATTRIBUTED TO PAPER PRODUCTION AND 22% TO PULP PRODUCTION.
3. THE BARK FLOW IS BASED ON DEBARKING 100% OF THE WOOD SUPPLY. FOR MILLS RECEIVING WOOD CHIPS, THE BARK FLOW WILL BE REDUCED IN PROPORTION TO THE AMOUNT OF CHIPS BEING RECEIVED. THEREFORE, TO SATISFY THE STEAM REQUIREMENTS ON THE "POWER PLANT ENERGY BALANCE", THE OIL AND COAL FLOWS MUST BE INCREASED ACCORDINGLY.
4. IN IDENTIFYING THE AIR EMISSIONS THE METHYL FRACTION HAS BEEN ABBREVIATED TO "R". THEREFORE:
    - RSH - METHYL MERCAPTAN
    - RSR - DIMETHYL SULFIDE
    - RSSR - DIMETHYL DISULFIDE
5. ABBREVIATIONS:
    - C. F. AND A. C. F. - ACTUAL CUBIC FEET AT FLOW CONDITIONS
    - S. D. C. F. - STANDARD DRY CUBIC FEET AT 70° F. AND 14.7 P.S.I.A.
    - EFF. - ANNUAL OPERATING EFFICIENCY AT THE CONTROL EQUIPMENT
6. ALL VALUES ARE ANNUAL AVERAGES.
7. ALL FUEL ANALYSES ARE ON AN "AS FIRED" BASIS.

KRAFT FLOW DIAGRAM Nº 2

CONTINUOUS DIGESTER, WEAK LIQUOR OXIDATION, DIRECT CONTACT EVAPORATOR, WITH BLEACH PLANT AND LIME KILN HIGH COLLECTION EFFICIENCY, CHLORINATION OF NON-CONDENSIBLE GASES

EXHIBIT NO.

SYSTEMS ANALYSIS STUDY OF EMISSIONS CONTROL IN THE WOOD PULPING INDUSTRY

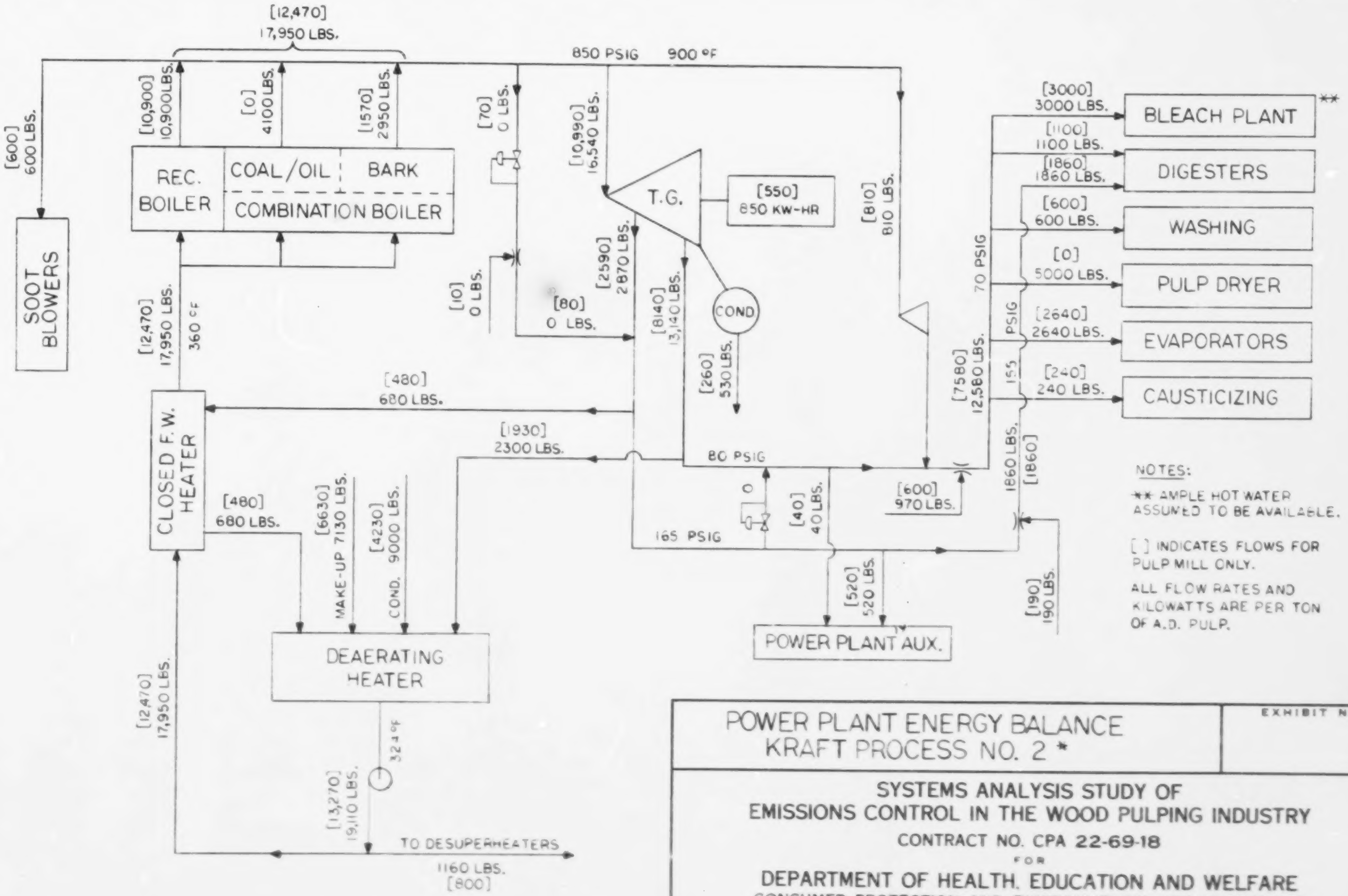
CONTRACT NO. CPA 22-69-18

FOR

DEPARTMENT OF HEALTH, EDUCATION AND WELFARE
CONSUMER PROTECTION AND ENVIRONMENTAL HEALTH SERVICE
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

ENVIRONMENTAL ENGINEERING, INC.
GAINESVILLE, FLORIDA

J. E. SIRRINE COMPANY, ENGINEERS
GREENVILLE, S. C.


[12,470]
17,950 LBS.
850 PSIG 900 °F
[600]
600 LBS.
SOOT BLOWERS
[10,900]
10,900 LBS.
[0]
4100 LBS.
[1570]
2950 LBS.
REC. BOILER
COAL/OIL
BARK
COMBINATION BOILER
[70]
0 LBS.
[10]
0 LBS.
[80]
0 LBS.
[10,990]
16,540 LBS.
[2590]
2870 LBS.
T.G.
[550]
850 KW-HR
[8140]
13,140 LBS.
COND
[260]
530 LBS.
[810]
810 LBS.
[12,470]
17,950 LBS.
360 °F
[480]
680 LBS.
[1930]
2300 LBS.
CLOSED F.W. HEATER
[480]
680 LBS.
[6630]
MAKE-UP 7130 LBS.
[4230]
COND. 9000 LBS.
80 PSIG
0
[40]
40 LBS.
165 PSIG
[600]
970 LBS.
[7580]
12,580 LBS.
70 PSIG
155 PSIG
1860 LBS.
[1860]
[520]
520 LBS.
POWER PLANT AUX.
[190]
190 LBS.
[3000]
3000 LBS.
BLEACH PLANT **
[1100]
1100 LBS.
DIGESTERS
[1860]
1860 LBS.
[600]
600 LBS.
WASHING
[0]
5000 LBS.
PULP DRYER
[2640]
2640 LBS.
EVAPORATORS
[240]
240 LBS.
CAUSTICIZING
DEAERATING HEATER
324 °F
[12,470]
17,950 LBS.
[13,270]
19,110 LBS.
TO DESUPERHEATERS
1160 LBS.
[800]

NOTES:

** AMPLE HOT WATER ASSUMED TO BE AVAILABLE.

[ ] INDICATES FLOWS FOR PULP MILL ONLY.

ALL FLOW RATES AND KILOWATTS ARE PER TON OF A.D. PULP.

POWER PLANT ENERGY BALANCE
KRAFT PROCESS NO. 2 *

EXHIBIT NO.

SYSTEMS ANALYSIS STUDY OF
EMISSIONS CONTROL IN THE WOOD PULPING INDUSTRY
CONTRACT NO. CPA 22-69-18
FOR
DEPARTMENT OF HEALTH, EDUCATION AND WELFARE
CONSUMER PROTECTION AND ENVIRONMENTAL HEALTH SERVICE
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

ENVIRONMENTAL ENGINEERING, INC.
GAINESVILLE, FLORIDA

J. E. SIRRINE COMPANY, ENGINEERS
GREENVILLE, S. C.

* CONTINUOUS DIGESTERS, WEAK BLACK LIQUOR OXIDATION, DIRECT CONTACT EVAPORATORS, BLEACHING, HIGH EFFICIENCY LIME KILN COLLECTION.

KRAFT DIAGRAM NO. 3

TYPICAL LOCATION

Western United States.

TYPICAL AGE OF EQUIPMENT

Ten to 15 years with upgraded emissions control equipment.

GENERAL

This flow diagram, based on production of unbleached softwood kraft pulp, depicts a mill designed with emphasis on odor abatement. It includes incineration in the lime kiln of the noncondensible gases, from the batch digesters and multiple-effect evaporators. There are perhaps 4 to 6 mills in the U. S. which are illustrated by this flow diagram.

EMISSIONS

The following emissions have been selected as representative, based on the range of emissions presented in Chapter 4.

POUNDS PER AIR DRY TON OF UNBLEACHED PULP

Location	H_2S	RSH, RSR, RSSR	SO_2	Particulate
Accumulator	0.10	3.35	0	0
Washers and Screens	0.02	0.20	0	0
BLO Tower	0.02	0.25	0	0
M. E. Evaporators	0.01	0.4	0	0
R. Boiler and D. C. Evaporator	5.0	Trace	4.6	106
After Precipitator	5.0	Trace	4.6	8.50

KRAFT DIAGRAM NO. 3
(Continued)

EMISSIONS - Continued

Location	H_2S	RSH, RSR, RSSR	SO_2	Particulate
After Scrubber	5.0	Trace	4.6	1.7
Smelt Dissolving Tank	0.02	0.02	0	1.0
Slaker	0	0	0	Unknown
Lime Kiln	1.00	0.6	Trace	51
Lime Kiln Scrubber	0.2	0.30	Trace	0.5
Turpentine Condenser	0.01	0.40	0	0

ASSUMPTIONS

The following assumptions have been made in developing the flow diagram:

A. PULP MILL

1. Pulp Yield = 47 Percent
2. Cooking liquor charge per air dry ton of brown stock equals 6,335 pounds of which 900 pounds are chemical solids. Sulfidity equals 25 percent.
3. It was assumed that there is no soap skimming system since only small quantities of soap are produced.
4. Weak liquor oxidation system ahead of the black liquor evaporators. Oxidation efficiency was assumed at 90 percent.

KRAFT DIAGRAM NO. 3
(Continued)

B. RECOVERY

1. Unit operating at 15 percent excess air at economizer outlet.

2. Particulate matter of 8 grains per SDCF leaving the economizer.

3. Efficiency of 50 percent on particulate removal in the direct contact evaporators.

4. Steam shatter jets utilized on smelt spouts.

5. Design efficiency of 95 percent for precipitator with an annual average operating efficiency of 92 percent.

6. Efficiency of 80 percent on particulate removal in the scrubber.

C. CAUSTICIZING

1. Lime kiln scrubber efficiency assumed at 99 percent on lime and 80 percent efficiency on soda.

D. POWER PLANT

1. Excess air: 30 percent for bark, 10 percent for oil and 20 percent for coal.

2. Bark burning equipment is a low set spreader stoker.

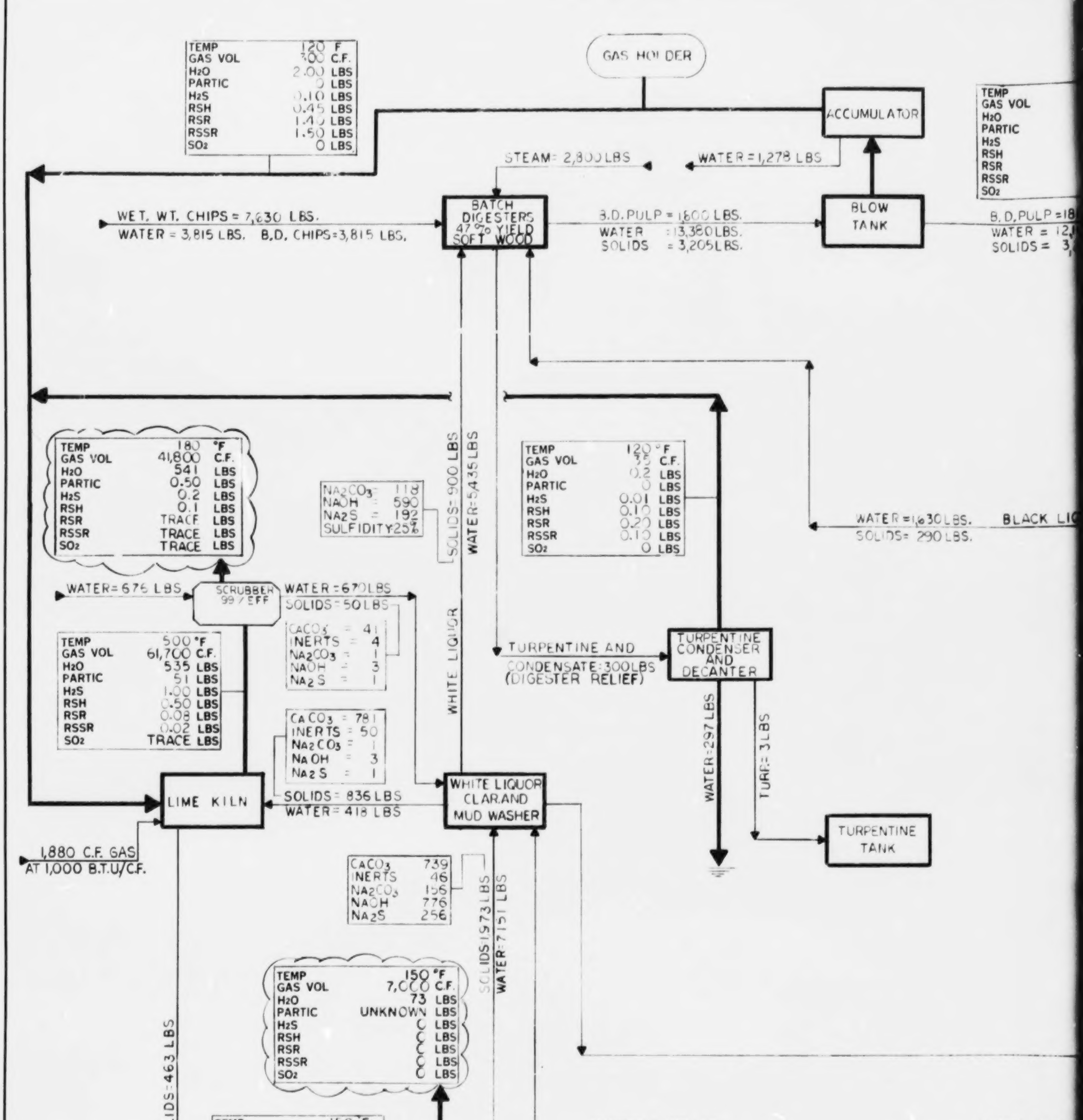
3. Reinjection from dust collector: 50 percent for bark and 0 percent for coal.

4. Design efficiencies for dust collector: 82 percent on bark and 96 percent on coal with annual operating efficiencies of 80 percent and 95 percent, respectively.

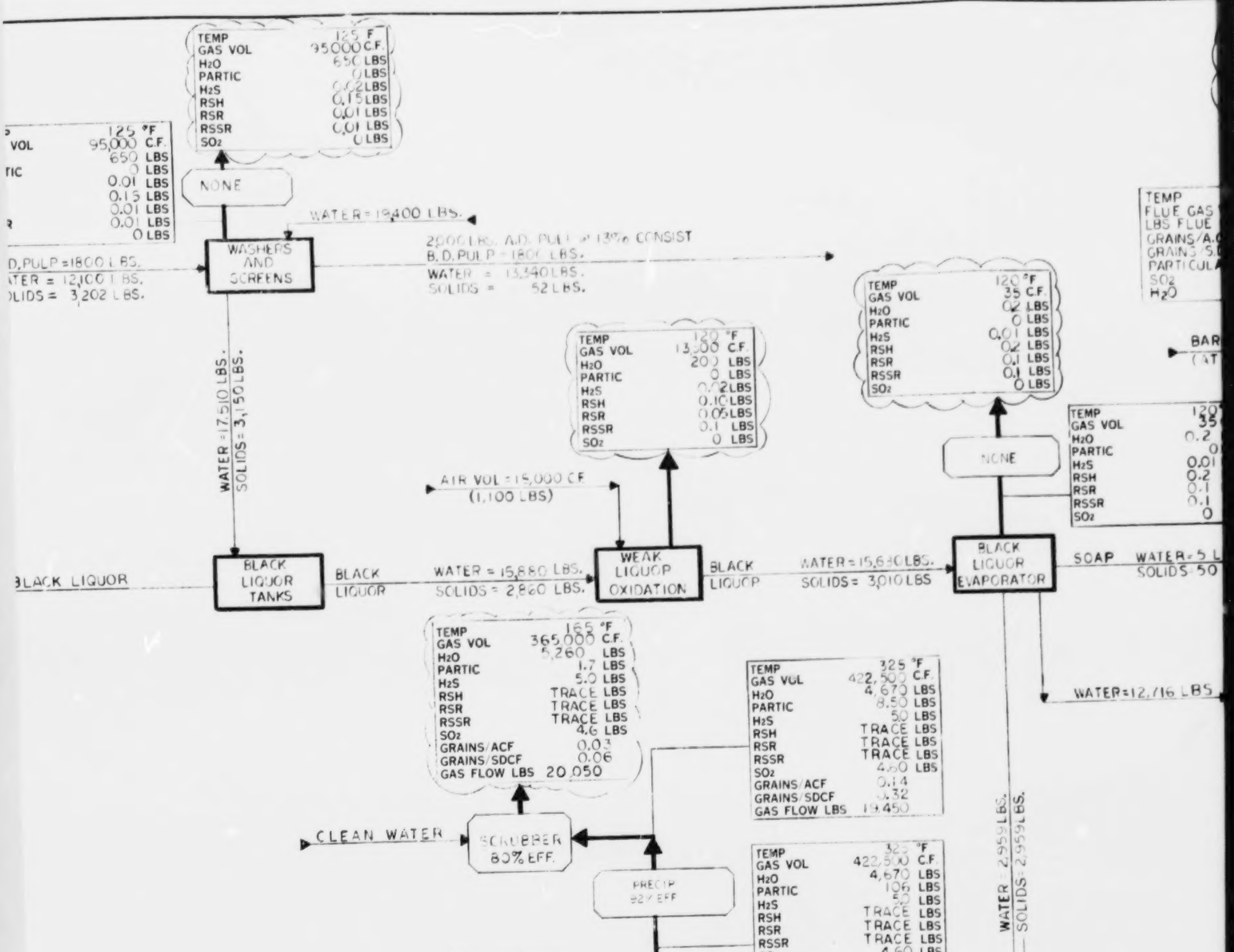
5. Bark is burned 24 hours per day at a controlled rate.

6. Unburned combustible in refuse leaving the dust collector is 40 percent.

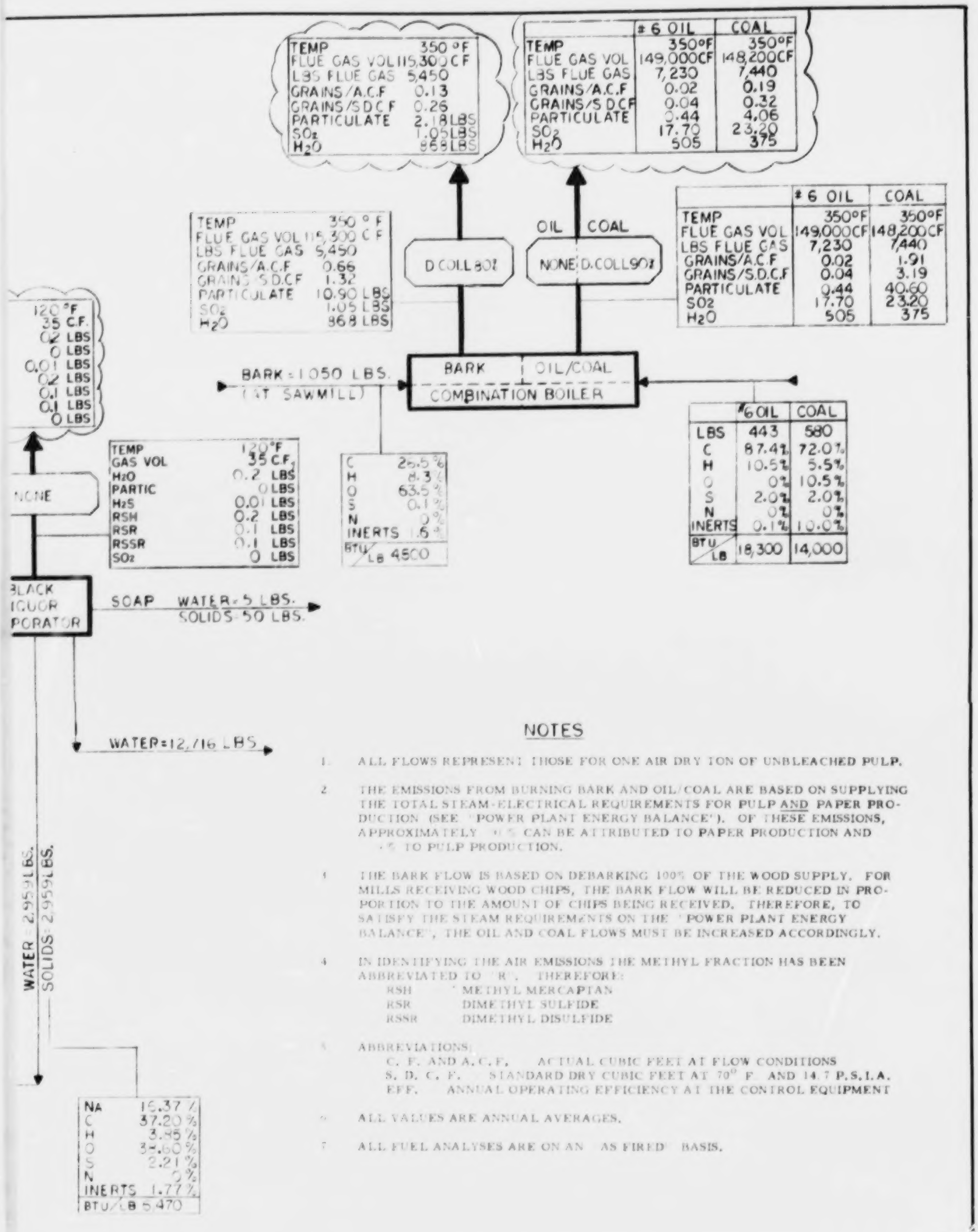
7. Coal firing based on pulverized coal.


GAS HOLDER
ACCUMULATOR
BLOW TANK
BATCH DIGESTERS 47% YIELD SOFT WOOD
TEMP 120 F
GAS VOL 300 C.F.
H2O 2.00 LBS
PARTIC 0 LBS
H2S 0.10 LBS
RSH 0.45 LBS
RSR 1.40 LBS
RSSR 1.50 LBS
SO2 0 LBS
STEAM= 2,800 LBS
WATER = 1,278 LBS
WET. WT. CHIPS = 7,630 LBS.
WATER = 3,815 LBS. B.D. CHIPS = 3,815 LBS.
B.D. PULP = 1,600 LBS.
WATER = 13,380 LBS.
SOLIDS = 3,205 LBS.
B.D. PULP = 18
WATER = 12,
SOLIDS = 3,
TEMP
GAS VOL
H2O
PARTIC
H2S
RSH
RSR
RSSR
SO2
TEMP 180 °F
GAS VOL 41,800 C.F.
H2O 541 LBS
PARTIC 0.50 LBS
H2S 0.2 LBS
RSH 0.1 LBS
RSR TRACE LBS
RSSR TRACE LBS
SO2 TRACE LBS
NA2CO3 = 118
NAOH = 590
NA2S = 192
SULFIDITY 25%
SOLIDS = 900 LBS
WATER = 5,435 LBS
WHITE LIQUOR
TEMP 120 °F
GAS VOL 35 C.F.
H2O 0.2 LBS
PARTIC 0 LBS
H2S 0.01 LBS
RSH 0.10 LBS
RSR 0.20 LBS
RSSR 0.10 LBS
SO2 0 LBS
WATER = 1,630 LBS.
SOLIDS = 290 LBS.
BLACK LI
WATER = 676 LBS
SCRUBBER 99% EFF
WATER = 670 LBS
SOLIDS = 50 LBS
CACO3 = 41
INERTS = 4
NA2CO3 = 1
NAOH = 3
NA2S = 1
TEMP 500 °F
GAS VOL 61,700 C.F.
H2O 535 LBS
PARTIC 51 LBS
H2S 1.00 LBS
RSH 0.50 LBS
RSR 0.08 LBS
RSSR 0.02 LBS
SO2 TRACE LBS
TURPENTINE AND CONDENSATE: 300 LBS (DIGESTER RELIEF)
TURPENTINE CONDENSER AND DECANTER
WATER = 297 LBS
TURP. = 3 LBS
TURPENTINE TANK
CACO3 = 781
INERTS = 50
NA2CO3 = 1
NAOH = 3
NA2S = 1
LIME KILN
SOLIDS = 836 LBS
WATER = 418 LBS
WHITE LIQUOR CLAR. AND MUD WASHER
1,880 C.F. GAS AT 1,000 B.T.U./C.F.
CACO3 739
INERTS 46
NA2CO3 156
NAOH 776
NA2S 256
SOLIDS 1973 LBS
WATER = 7151 LBS
TEMP 150 °F
GAS VOL 7,000 C.F.
H2O 73 LBS
PARTIC UNKNOWN LBS
H2S 0 LBS
RSH 0 LBS
RSR 0 LBS
RSSR 0 LBS
SO2 0 LBS
SOLIDS = 463 LBS
CAO = 413
INERTS = 50
TEMP 150 °F
GAS VOL 7,000 C.F.
H2O 73 LBS
PARTIC UNKNOWN LBS
H2S 0 LBS
RSH 0 LBS
RSR 0 LBS
RSSR 0 LBS
SO2 0 LBS
NONE
WATER = 4,149 LBS
WATER = 6,120
WATER = 1971 LBS
SLAKER
WATER = 7,284 LBS
SOLIDS = 1,379 LBS.
GREEN LIQUOR CLAR. AND DREGS WASHER
GREEN
SOLIDS = 25 LBS
CAO 22.5
INERTS 2.5
CAO 12.5
INERTS = 2.5
SOLIDS = 15 LB
NA2CO3 = 941
NAOH = 186
NA2S = 252
INERTS = 15.0
NA2CO3 = 0.9
NAOH = 0.1
NA2S = 0.1
SOLIDS 16 LB
WATER = 134 LB

A


TEMP 125 °F
GAS VOL 95,000 C.F.
650 LBS
TIC 0 LBS
0.01 LBS
0.15 LBS
0.01 LBS
R 0.01 LBS
0 LBS
TEMP 125 F
GAS VOL 95000 C.F.
H2O 650 LBS
PARTIC 0 LBS
H2S 0.02 LBS
RSH 0.15 LBS
RSR 0.01 LBS
RSSR 0.01 LBS
SO2 0 LBS
NONE
WATER = 19,400 LBS.
D.PULP = 1800 LBS.
ATER = 12,100 LBS.
OLIDS = 3,202 LBS.
WASHERS AND SCREENS
2000 LBS. A.D. PULP @ 13% CONSIST
B.D. PULP = 1800 LBS.
WATER = 13,340 LBS.
SOLIDS = 52 LBS.
WATER = 17,510 LBS.
SOLIDS = 3,150 LBS.
BLACK LIQUOR
BLACK LIQUOR TANKS
BLACK LIQUOR
WATER = 15,880 LBS.
SOLIDS = 2,860 LBS.
TEMP 120 °F
GAS VOL 13,000 C.F.
H2O 200 LBS
PARTIC 0 LBS
H2S 0.02 LBS
RSH 0.10 LBS
RSR 0.05 LBS
RSSR 0.1 LBS
SO2 0 LBS
AIR VOL = 15,000 C.F.
(1,100 LBS)
WEAK LIQUOR OXIDATION
BLACK LIQUOR
WATER = 15,630 LBS.
SOLIDS = 3,010 LBS
TEMP 120 °F
GAS VOL 35 C.F.
H2O 0.2 LBS
PARTIC 0 LBS
H2S 0.01 LBS
RSH 0.2 LBS
RSR 0.1 LBS
RSSR 0.1 LBS
SO2 0 LBS
NONE
TEMP 120
GAS VOL 35
H2O 0.2
PARTIC 0
H2S 0.01
RSH 0.2
RSR 0.1
RSSR 0.1
SO2 0
BLACK LIQUOR EVAPORATOR
SOAP
WATER = 5 L
SOLIDS 50
WATER = 12,716 LBS
TEMP
FLUE GAS
LBS FLUE
GRAINS/A.C
GRAINS/S.D
PARTICULA
SO2
H2O
BAR
(AT
TEMP 165 °F
GAS VOL 365,000 C.F.
H2O 5,260 LBS
PARTIC 1.7 LBS
H2S 5.0 LBS
RSH TRACE LBS
RSR TRACE LBS
RSSR TRACE LBS
SO2 4.6 LBS
GRAINS/ACF 0.03
GRAINS/SDCF 0.06
GAS FLOW LBS 20,050
TEMP 325 °F
GAS VOL 422,500 C.F.
H2O 4,670 LBS
PARTIC 8.50 LBS
H2S 5.0 LBS
RSH TRACE LBS
RSR TRACE LBS
RSSR TRACE LBS
SO2 4.60 LBS
GRAINS/ACF 0.14
GRAINS/SDCF 0.32
GAS FLOW LBS 19,450
CLEAN WATER
SCRUBBER 80% EFF.
PRECIP 92% EFF
TEMP 325 °F
GAS VOL 422,500 C.F.
H2O 4,670 LBS
PARTIC 106 LBS
H2S 5.0 LBS
RSH TRACE LBS
RSR TRACE LBS
RSSR TRACE LBS
SO2 4.60 LBS
GRAINS/ACF 1.80
GRAINS/SDCF 4.00
GAS FLOW LBS 19,450
WATER = 2,959 LBS.
SOLIDS = 2,959 LBS.
(SALT CAKE) Na2SO4
RECOVERY BOILER AND CONTACT EVAPORATOR
NA 16.37 %
C 37.20 %
H 3.85 %
O 38.60 %
S 2.21 %
N 0 %
INERTS 1.77 %
BTU/LB 5,470
TEMP 190 °F
GAS VOL 44,500 C.F.
H2O 690 LBS
PARTIC 1.00 LBS
H2S 0.02 LBS
RSH 0.01 LBS
RSR 0.01 LBS
RSSR 0 LBS
SO2 0 LBS
GRAINS/ACF 0.20
GRAINS/SDCF 0.30
GAS FLOW LBS 2,320
SOLIDS = 1113 LBS
INERTS = 15
Na2CO3 = 906
Na2S = 192
MESH PAD 75% EFF
TEMP 200 °F
GAS VOL 45,000 C.F.
H2O 670 LBS
PARTIC 4.00 LBS
H2S 0.04 LBS
RSH 0.04 LBS
RSR 0.02 LBS
RSSR 0 LBS
SO2 0 LBS
GRAINS/ACF 0.60
GRAINS/SDCF 1.30
GAS FLOW LBS 2,320
R = 6,120 LBS.
Na2CO3 = 37
NaOH = 186
Na2S = 64
SOLIDS 287 LBS.
WATER 6,117 LBS.
SMELT DISSOLVING TANK
GREEN LIQUOR
WATER = 7,398 LBS.
SOLIDS = 1,415 LBS.
INERTS = 15
Na2CO3 = 956
NaOH = 189
Na2S = 256
Na2CO3 = 14
NaOH = 2
Na2S = 4
SOLIDS = 20 LBS
WATER = 1951 LBS
LEGEND
- PROCESS EQUIPMENT
- CONTROL EQUIPMENT
- AIR EMISSIONS
BATCH
NO BL
INCIN
ENV

B


TEMP 350 °F
FLUE GAS VOL 115,300 CF
LBS FLUE GAS 5,450
GRAINS/A.C.F 0.13
GRAINS/S D C F 0.26
PARTICULATE 2.18 LBS
SO_2 1.05 LBS
H_2O 868 LBS
# 6 OIL
COAL
TEMP 350°F 350°F
FLUE GAS VOL 149,000CF 148,200CF
LBS FLUE GAS 7,230 7,440
GRAINS/A.C.F 0.02 0.19
GRAINS/S D CF 0.04 0.32
PARTICULATE 0.44 4.06
SO_2 17.70 23.20
H_2O 505 375
TEMP 350 °F
FLUE GAS VOL 115,300 C F
LBS FLUE GAS 5,450
GRAINS/A.C.F 0.66
GRAINS/S.D.C.F 1.32
PARTICULATE 10.90 LBS
SO_2 1.05 LBS
H_2O 868 LBS
D COLL 80%
OIL
COAL
NONE
D.COLL 90%
# 6 OIL
COAL
TEMP 350°F 350°F
FLUE GAS VOL 149,000CF 148,200CF
LBS FLUE GAS 7,230 7,440
GRAINS/A.C.F 0.02 1.91
GRAINS/S.D.C.F 0.04 3.19
PARTICULATE 0.44 40.60
SO2 17.70 23.20
H2O 505 375
120 °F
35 C.F.
0.2 LBS
0 LBS
0.01 LBS
0.2 LBS
0.1 LBS
0.1 LBS
0 LBS
BARK = 1050 LBS.
(AT SAWMILL)
BARK
OIL/COAL
COMBINATION BOILER
#6 OIL
COAL
LBS 443 580
C 87.4% 72.0%
H 10.5% 5.5%
O 0% 10.5%
S 2.0% 2.0%
N 0% 0%
INERTS 0.1% 10.0%
BTU/LB 18,300 14,000
TEMP 120°F
GAS VOL 35 C.F.
H2O 0.2 LBS
PARTIC 0 LBS
H2S 0.01 LBS
RSH 0.2 LBS
RSR 0.1 LBS
RSSR 0.1 LBS
SO2 0 LBS
C 25.5%
H 8.3%
O 63.5%
S 0.1%
N 0%
INERTS 1.6%
BTU/LB 4500
NONE
BLACK
LIQUOR
PORATOR
SOAP
WATER = 5 LBS.
SOLIDS 50 LBS.
WATER = 12,716 LBS
WATER = 2,959 LBS.
SOLIDS = 2,959 LBS.
NA 16.37%
C 37.20%
H 3.85%
O 35.60%
S 2.21%
N 0%
INERTS 1.77%
BTU/LB 5,470

NOTES

1. ALL FLOWS REPRESENT THOSE FOR ONE AIR DRY TON OF UNBLEACHED PULP.
2. THE EMISSIONS FROM BURNING BARK AND OIL/COAL ARE BASED ON SUPPLYING THE TOTAL STEAM-ELECTRICAL REQUIREMENTS FOR PULP AND PAPER PRODUCTION (SEE "POWER PLANT ENERGY BALANCE"). OF THESE EMISSIONS, APPROXIMATELY [illegible] CAN BE ATTRIBUTED TO PAPER PRODUCTION AND [illegible] TO PULP PRODUCTION.
3. THE BARK FLOW IS BASED ON DEBARKING 100% OF THE WOOD SUPPLY. FOR MILLS RECEIVING WOOD CHIPS, THE BARK FLOW WILL BE REDUCED IN PROPORTION TO THE AMOUNT OF CHIPS BEING RECEIVED. THEREFORE, TO SATISFY THE STEAM REQUIREMENTS ON THE "POWER PLANT ENERGY BALANCE", THE OIL AND COAL FLOWS MUST BE INCREASED ACCORDINGLY.
4. IN IDENTIFYING THE AIR EMISSIONS THE METHYL FRACTION HAS BEEN ABBREVIATED TO "R". THEREFORE:
   - RSH = METHYL MERCAPTAN
   - RSR DIMETHYL SULFIDE
   - RSSR DIMETHYL DISULFIDE
5. ABBREVIATIONS:
   - C. F. AND A. C. F. ACTUAL CUBIC FEET AT FLOW CONDITIONS
   - S. D. C. F. STANDARD DRY CUBIC FEET AT 70° F AND 14.7 P.S.I.A.
   - EFF. ANNUAL OPERATING EFFICIENCY AT THE CONTROL EQUIPMENT
6. ALL VALUES ARE ANNUAL AVERAGES.
7. ALL FUEL ANALYSES ARE ON AN "AS FIRED" BASIS.

KRAFT FLOW DIAGRAM Nº 3

BATCH DIGESTER, WEAK LIQUOR OXIDATION, DIRECT CONTACT EVAPORATOR, NO BLEACHING, LIME KILN [illegible] COLLECTION EFFICIENCY, INCINERATION OF NON-CONDENSIBLE GASES

EXHIBIT NO.

SYSTEMS ANALYSIS STUDY OF EMISSIONS CONTROL IN THE WOOD PULPING INDUSTRY

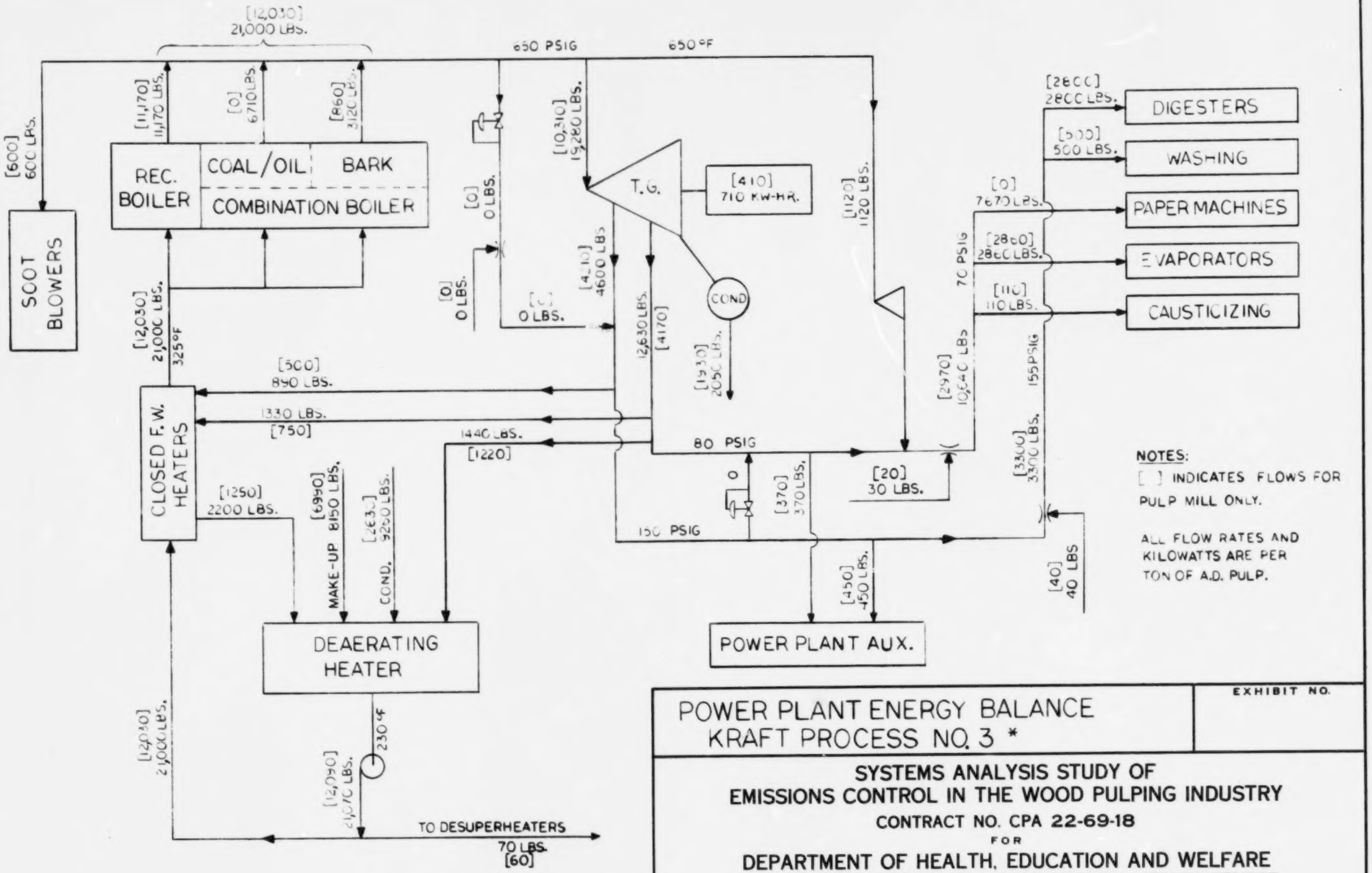
CONTRACT NO. CPA 22-69-18
FOR
DEPARTMENT OF HEALTH, EDUCATION AND WELFARE
CONSUMER PROTECTION AND ENVIRONMENTAL HEALTH SERVICE
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

ENVIRONMENTAL ENGINEERING, INC.
GAINESVILLE, FLORIDA

J. E. SIRRINE COMPANY, ENGINEERS
GREENVILLE, S. C.

GEND
ESS EQUIPMENT
ROL EQUIPMENT
EMISSIONS

C


[12,030]
21,000 LBS.
650 PSIG
650 °F
[600]
600 LBS.
SOOT BLOWERS
[11,170]
11,170 LBS.
[0]
6710 LBS.
[860]
3120 LBS.
REC. BOILER
COAL/OIL
BARK
COMBINATION BOILER
[12,030]
21,000 LBS.
325°F
[0]
0 LBS.
[0]
0 LBS.
[0]
0 LBS.
[10,310]
19,280 LBS.
T.G.
[410]
710 KW-HR.
[4,110]
4600 LBS.
12,630 LBS.
[4170]
COND
[1930]
2050 LBS.
[1120]
1120 LBS.
[500]
890 LBS.
1330 LBS.
[750]
1440 LBS.
[1220]
80 PSIG
[2970]
10,640 LBS
70 PSIG
[2800]
2800 LBS.
DIGESTERS
[500]
500 LBS.
WASHING
[0]
7670 LBS.
PAPER MACHINES
[2860]
2860 LBS.
EVAPORATORS
[110]
110 LBS.
CAUSTICIZING
155 PSIG
[3300]
3300 LBS.
[20]
30 LBS.
0
[370]
370 LBS.
150 PSIG
[450]
450 LBS.
POWER PLANT AUX.
[40]
40 LBS
CLOSED F.W. HEATERS
[1250]
2200 LBS.
MAKE-UP 8150 LBS.
[6990]
COND. 9250 LBS.
[2630]
DEAERATING HEATER
230 °F
[12,090]
21,070 LBS.
[12,030]
21,000 LBS.
TO DESUPERHEATERS
70 LBS.
[60]

NOTES:

[ ] INDICATES FLOWS FOR PULP MILL ONLY.

ALL FLOW RATES AND KILOWATTS ARE PER TON OF A.D. PULP.

* BATCH DIGESTERS, BASE STOCK, WEAK LIQUOR OXIDATION, NO BLEACHING, DIRECT CONTACT EVAPORATORS, HIGH LIME KILN COLLECTION EFFICIENCY.

POWER PLANT ENERGY BALANCE
KRAFT PROCESS NO. 3 *

EXHIBIT NO.

SYSTEMS ANALYSIS STUDY OF
EMISSIONS CONTROL IN THE WOOD PULPING INDUSTRY
CONTRACT NO. CPA 22-69-18
FOR
DEPARTMENT OF HEALTH, EDUCATION AND WELFARE
CONSUMER PROTECTION AND ENVIRONMENTAL HEALTH SERVICE
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

ENVIRONMENTAL ENGINEERING, INC.
GAINESVILLE, FLORIDA

J. E. SIRRINE COMPANY, *ENGINEERS*
GREENVILLE, S. C.

KRAFT DIAGRAM NO. 4

TYPICAL LOCATION

Southern United States.

TYPICAL AGE OF EQUIPMENT

Under five years.

GENERAL

This flow diagram is based on production of kraft softwood paper stock. It depicts a new mill designed to provide some odor abatement, consisting of a concentrated liquor oxidation system, a scrubber in the noncondensible gases from the black liquor evaporators and a mesh pad in the dissolving tank vent. A comparison of emissions from the lime kiln is shown when using clean water make-up in the causticizing plant versus the use of evaporator combined condensate. There are perhaps 10 to 20 mills in the U. S. that are illustrated by this flow diagram.

EMISSIONS

The following emissions have been selected as representative, based on the range of emissions presented in Chapter 4.

POUNDS PER AIR DRY TON OF UNBLEACHED PULP

LOCATION	H_2S	RSH, RSR, RSSR	SO_2	Particulate
Washers and Screens	0.02	0.24	0	0
M. E. Evaporators	0.40	0.41	0	0
BLO Tower	0.02	0.31	0	0
R. Boiler and D. C. Evaporators	1.5	0.12	4.4	106
After Precipitators	1.5	0.12	4.4	3.18
Smelt Dissolving Tank	0.02	0.02	0	1.0
Slaker	0	0	0	Unknown

KRAFT DIAGRAM NO. 4
(Continued)

EMISSIONS - Continued

POUNDS PER AIR DRY TON OF UNBLEACHED PULP

LOCATION	H_2S	RSH, RSR, RSSR	SO_2	Particulate
Lime Kiln	1.0	0.5	Trace	25
Lime Kiln Scrubber	0.2	0.10	Trace	0.2
Turpentine Condenser	0.01	0.50	0	0

ASSUMPTIONS

The following assumptions have been made in developing the flow diagram:

A. PULP MILL

1. Pulp Yield = 47 Percent.
2. Cooking liquor charge per air dry ton of brown stock equals 6,189 pounds of which 889 pounds are chemical solids. Sulfidity equals 26 percent.
3. Assumed concentrated black liquor oxidation at approximately 99 percent efficiency.

B. RECOVERY

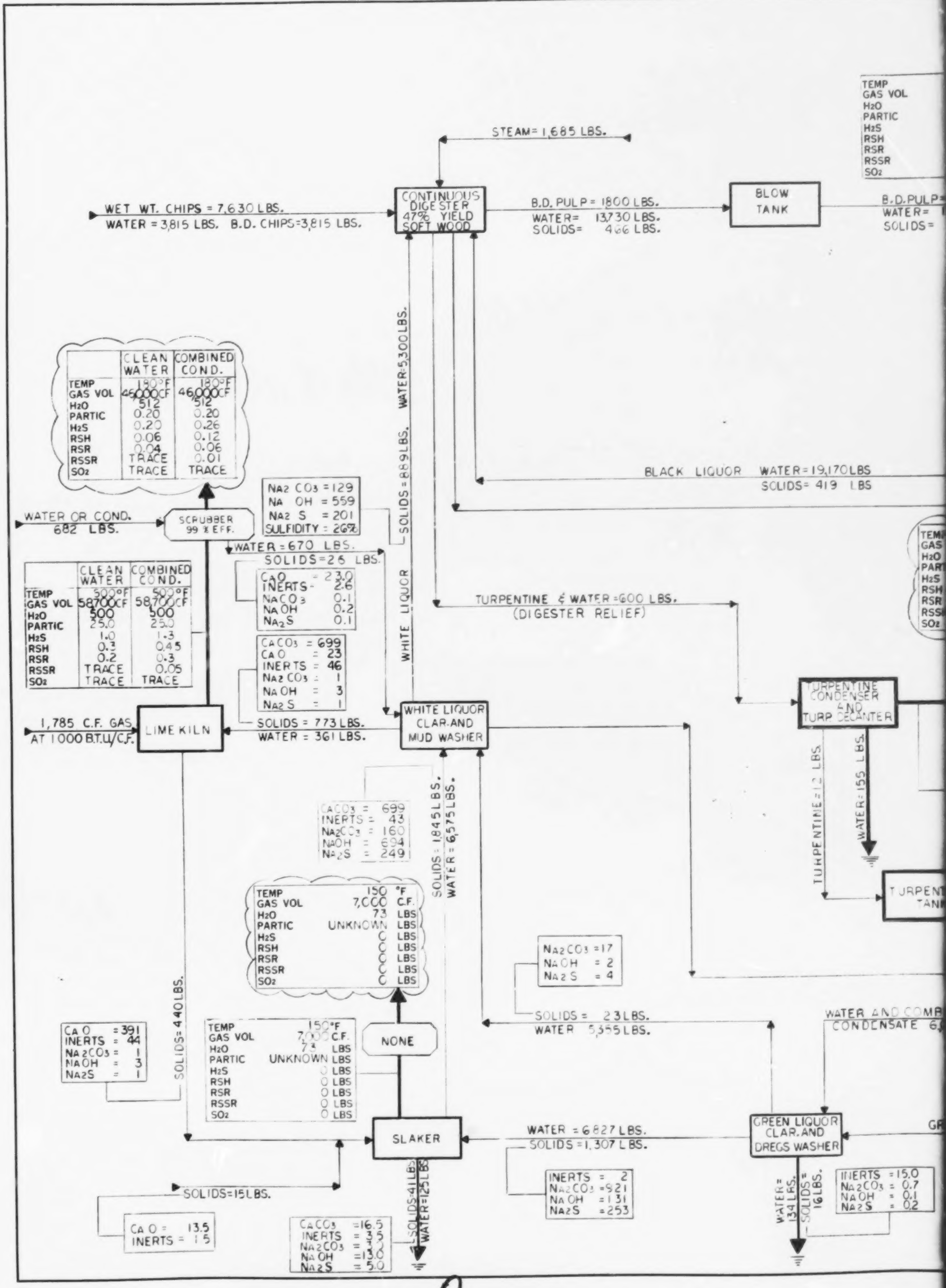
1. Unit operating at 15 percent excess air at economizer outlet.
2. Particulate matter of 8 grains per SDCF leaving the economizer.
3. Efficiency of 50 percent on particulate removal in the direct contact evaporators.
4. Steam shatter jets utilized on smelt spouts.
5. Design efficiency of 99 percent for precipitator with an annual average operating efficiency of 97 percent.

C. CAUSTICIZING

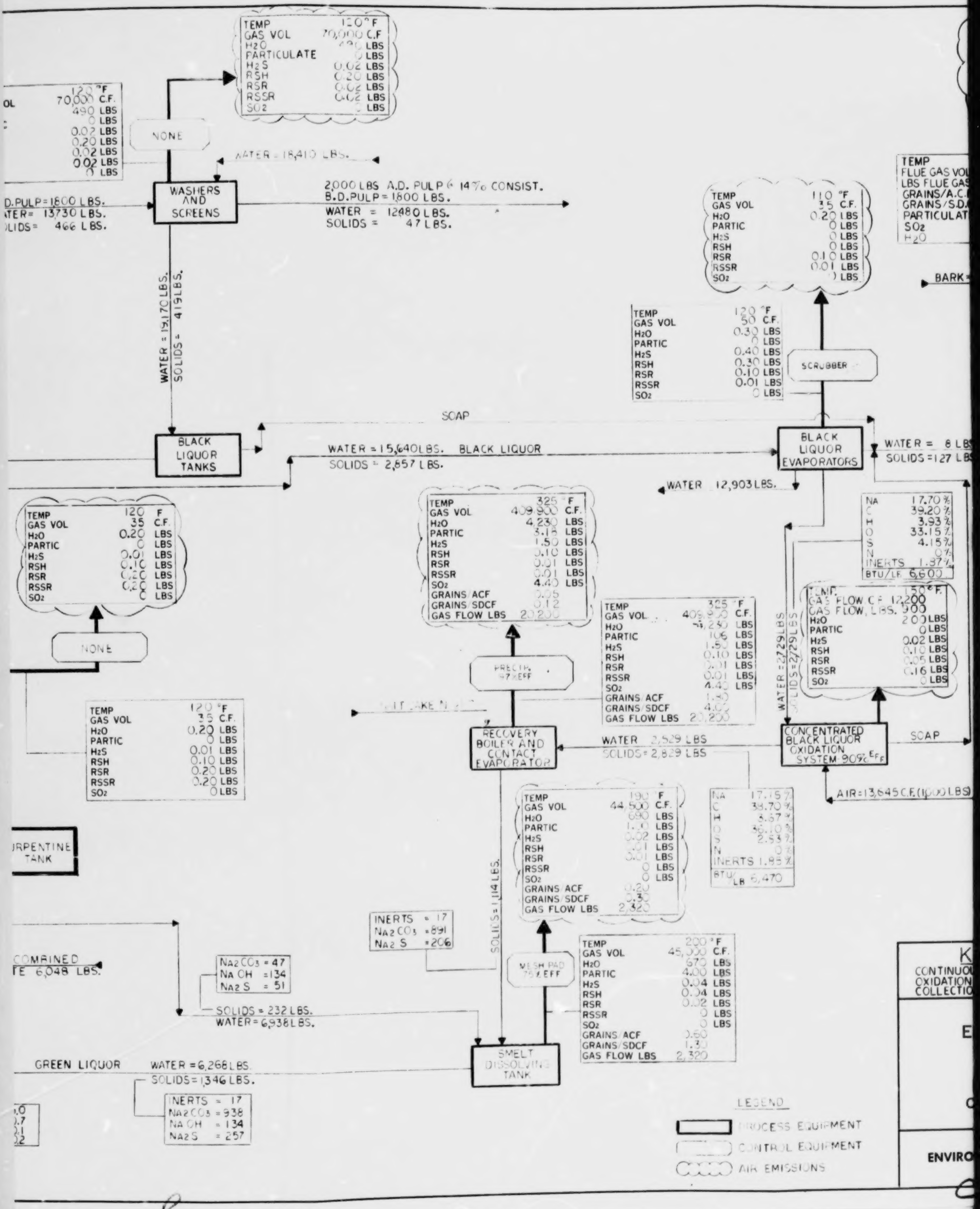
1. Lime kiln scrubber efficiency assumed at 99 percent on lime and 80 percent efficiency on soda.

D. POWER PLANT

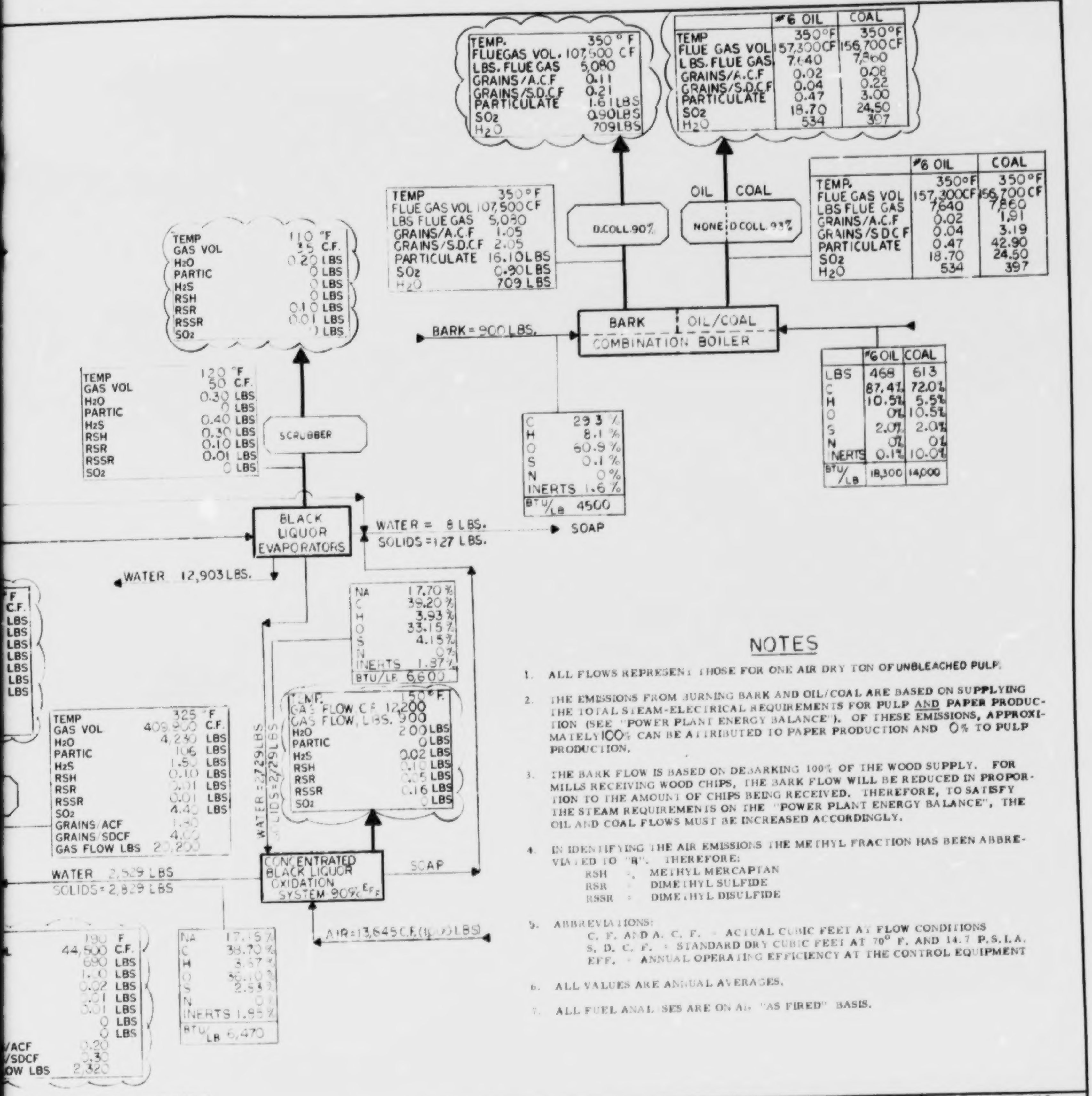
1. Excess air: 30 percent for bark, 10 percent for oil and 20 percent for coal.
2. Bark burning equipment is a low set spreader stoker.
3. Reinjection from dust collector: 50 percent for bark and 0 percent for coal.
4. Design efficiencies for dust collector: 92 percent on bark and 95 percent on coal with annual operating efficiencies of 90 percent and 93 percent, respectively.
5. Bark is burned 24 hours per day at a controlled rate.
6. Unburned combustible in refuse leaving the dust collector is 40 percent.
7. Coal firing based on pulverized coal.


STEAM= 1,685 LBS.
WET WT. CHIPS = 7,630 LBS.
WATER = 3,815 LBS. B.D. CHIPS=3,815 LBS.
CONTINUOUS DIGESTER 47% YIELD SOFT WOOD
B.D. PULP = 1800 LBS.
WATER= 13,730 LBS.
SOLIDS= 466 LBS.
BLOW TANK
B.D.PULP=
WATER=
SOLIDS=
TEMP
GAS VOL
H2O
PARTIC
H2S
RSH
RSR
RSSR
SO2
SOLIDS=889 LBS. WATER=5,300 LBS.
WHITE LIQUOR
CLEAN WATER | COMBINED COND.
TEMP 180°F 180°F
GAS VOL 46,000CF 46,000CF
H2O 512 512
PARTIC 0.20 0.20
H2S 0.20 0.26
RSH 0.06 0.12
RSR 0.04 0.06
RSSR TRACE 0.01
SO2 TRACE TRACE
BLACK LIQUOR WATER=19,170 LBS
SOLIDS= 419 LBS
NA2 CO3 =129
NA OH = 559
NA2 S = 201
SULFIDITY = 26%
WATER OR COND. 682 LBS.
SCRUBBER 99 % EFF.
WATER = 670 LBS.
SOLIDS=26 LBS.
CAO = 23.0
INERTS - 2.6
NACO3 0.1
NA OH 0.2
NA2S 0.1
TURPENTINE & WATER = 600 LBS. (DIGESTER RELIEF)
CLEAN WATER | COMBINED COND.
TEMP 500°F 500°F
GAS VOL 58700CF 58700CF
H2O 500 500
PARTIC 25.0 25.0
H2S 1.0 1.3
RSH 0.3 0.45
RSR 0.2 0.3
RSSR TRACE 0.05
SO2 TRACE TRACE
CACO3 = 699
CAO = 23
INERTS = 46
NA2 CO3 = 1
NA OH = 3
NA2 S = 1
TURPENTINE CONDENSER AND TURP DECANTER
1,785 C.F. GAS AT 1000 B.T.U/C.F.
LIME KILN
SOLIDS = 773 LBS.
WATER = 361 LBS.
WHITE LIQUOR CLAR. AND MUD WASHER
TURPENTINE=12 LBS.
WATER=155 LBS.
CACO3 = 699
INERTS = 43
NA2CO3 = 160
NAOH = 694
NA2S = 249
SOLIDS = 1,845 LBS.
WATER = 6,575 LBS.
TURPENT TANK
TEMP 150 °F
GAS VOL 7,000 C.F.
H2O 73 LBS
PARTIC UNKNOWN LBS
H2S 0 LBS
RSH 0 LBS
RSR 0 LBS
RSSR 0 LBS
SO2 0 LBS
NA2CO3 = 17
NAOH = 2
NA2S = 4
SOLIDS= 440 LBS.
SOLIDS = 23 LBS.
WATER 5,355 LBS.
WATER AND COMB CONDENSATE 6,
CA O = 391
INERTS = 44
NA2CO3 = 1
NAOH = 3
NA2S = 1
TEMP 150°F
GAS VOL 7,000 C.F.
H2O 73 LBS
PARTIC UNKNOWN LBS
H2S 0 LBS
RSH 0 LBS
RSR 0 LBS
RSSR 0 LBS
SO2 0 LBS
NONE
SLAKER
WATER = 6827 LBS.
SOLIDS = 1,307 LBS.
GREEN LIQUOR CLAR. AND DREGS WASHER
GR
SOLIDS=15 LBS.
SOLIDS=41 LBS.
WATER=125 LBS.
INERTS = 2
NA2CO3 = 921
NA OH = 131
NA2S = 253
WATER= 134 LBS.
SOLIDS= 16 LBS.
INERTS = 15.0
NA2CO3 = 0.7
NAOH = 0.1
NA2S = 0.2
CA O = 13.5
INERTS = 1.5
CACO3 = 16.5
INERTS = 3.5
NA2CO3 = 3.3
NA OH = 13.0
NA2S = 5.0

A


TEMP 120 °F
GAS VOL 70,000 C.F.
H2O 490 LBS
PARTICULATE 0 LBS
H2S 0.02 LBS
RSH 0.20 LBS
RSR 0.02 LBS
RSSR 0.02 LBS
SO2 0 LBS
120 °F
OL 70,000 C.F.
490 LBS
0 LBS
0.02 LBS
0.20 LBS
0.02 LBS
0.02 LBS
0 LBS
NONE
WATER = 18,410 LBS.
WASHERS AND SCREENS
D.PULP = 1800 LBS.
TER = 13730 LBS.
LIDS = 466 LBS.
2,000 LBS A.D. PULP @ 14% CONSIST.
B.D.PULP = 1800 LBS.
WATER = 12480 LBS.
SOLIDS = 47 LBS.
WATER = 19,170 LBS.
SOLIDS = 419 LBS.
SOAP
BLACK LIQUOR TANKS
WATER = 15,640 LBS. BLACK LIQUOR
SOLIDS = 2,857 LBS.
TEMP 110 °F
GAS VOL 35 C.F.
H2O 0.20 LBS
PARTIC 0 LBS
H2S 0 LBS
RSH 0 LBS
RSR 0.10 LBS
RSSR 0.01 LBS
SO2 0 LBS
TEMP 120 °F
GAS VOL 50 C.F.
H2O 0.30 LBS
PARTIC 0 LBS
H2S 0.40 LBS
RSH 0.30 LBS
RSR 0.10 LBS
RSSR 0.01 LBS
SO2 0 LBS
SCRUBBER
BLACK LIQUOR EVAPORATORS
WATER 12,903 LBS.
WATER = 8 LBS
SOLIDS = 127 LBS
TEMP
FLUE GAS VOL
LBS FLUE GAS
GRAINS/A.C.
GRAINS/S.D.
PARTICULAT
SO2
H2O
BARK =
NA 17.70 %
C 39.20 %
H 3.93 %
O 33.15 %
S 4.15 %
N 0 %
INERTS 1.87 %
BTU/LB 6,600
TEMP 150 °F
GAS FLOW C.F. 12,200
GAS FLOW, LBS. 900
H2O 200 LBS
PARTIC 0 LBS
H2S 0.02 LBS
RSH 0.10 LBS
RSR 0.05 LBS
RSSR 0.16 LBS
SO2 0 LBS
WATER = 2729 LBS
SOLIDS = 2729 LBS
CONCENTRATED BLACK LIQUOR OXIDATION SYSTEM 90% EFF
SOAP
AIR = 13,645 C.F. (1000 LBS)
TEMP 120 F
GAS VOL 35 C.F.
H2O 0.20 LBS
PARTIC 0 LBS
H2S 0.01 LBS
RSH 0.10 LBS
RSR 0.20 LBS
RSSR 0.20 LBS
SO2 0 LBS
NONE
TEMP 120 °F
GAS VOL 35 C.F.
H2O 0.20 LBS
PARTIC 0 LBS
H2S 0.01 LBS
RSH 0.10 LBS
RSR 0.20 LBS
RSSR 0.20 LBS
SO2 0 LBS
TEMP 325 °F
GAS VOL 409,900 C.F.
H2O 4,230 LBS
PARTIC 3.18 LBS
H2S 1.50 LBS
RSH 0.10 LBS
RSR 0.01 LBS
RSSR 0.01 LBS
SO2 4.40 LBS
GRAINS/ACF 0.05
GRAINS/SDCF 0.12
GAS FLOW LBS 20,200
PRECTR. 97% EFF
TEMP 325 °F
GAS VOL 409,900 C.F.
H2O 4,230 LBS
PARTIC 106 LBS
H2S 1.50 LBS
RSH 0.10 LBS
RSR 0.01 LBS
RSSR 0.01 LBS
SO2 4.40 LBS
GRAINS/ACF 1.80
GRAINS/SDCF 4.00
GAS FLOW LBS 20,200
…LT CAKE N…
RECOVERY BOILER AND CONTACT EVAPORATOR
WATER 2,529 LBS
SOLIDS = 2,829 LBS
NA 17.15 %
C 38.70 %
H 3.67 %
O 36.10 %
S 2.53 %
N 0 %
INERTS 1.85 %
BTU/LB 6,470
TEMP 190 °F
GAS VOL 44,500 C.F.
H2O 690 LBS
PARTIC 1.00 LBS
H2S 0.02 LBS
RSH 0.01 LBS
RSR 0.01 LBS
RSSR 0 LBS
SO2 0 LBS
GRAINS/ACF 0.20
GRAINS/SDCF 0.30
GAS FLOW LBS 2,320
URPENTINE TANK
SOLIDS = 1,114 LBS.
INERTS = 17
Na2CO3 = 891
Na2S = 206
MESH PAD 75% EFF
TEMP 200 °F
GAS VOL 45,000 C.F.
H2O 670 LBS
PARTIC 4.00 LBS
H2S 0.04 LBS
RSH 0.04 LBS
RSR 0.02 LBS
RSSR 0 LBS
SO2 0 LBS
GRAINS/ACF 0.60
GRAINS/SDCF 1.30
GAS FLOW LBS 2,320
COMBINED
TE 6,048 LBS.
Na2CO3 = 47
NaOH = 134
Na2S = 51
SOLIDS = 232 LBS.
WATER = 6,938 LBS.
SMELT DISSOLVING TANK
GREEN LIQUOR
WATER = 6,268 LBS.
SOLIDS = 1,346 LBS.
INERTS = 17
Na2CO3 = 938
NaOH = 134
Na2S = 257
.0
.7
.1
.2
LEGEND
PROCESS EQUIPMENT
CONTROL EQUIPMENT
AIR EMISSIONS
K
CONTINUO
OXIDATION
COLLECTIO
E
C
ENVIRO

B


TEMP. 350 °F
FLUEGAS VOL. 107,500 CF
LBS. FLUE GAS 5,080
GRAINS/A.C.F 0.11
GRAINS/S.D.C.F 0.21
PARTICULATE 1.61 LBS
SO2 0.90 LBS
H2O 709 LBS
#6 OIL
COAL
TEMP 350°F 350°F
FLUE GAS VOL 157,300 CF 156,700 CF
LBS. FLUE GAS 7,640 7,860
GRAINS/A.C.F 0.02 0.08
GRAINS/S.D.C.F 0.04 0.22
PARTICULATE 0.47 3.00
SO2 18.70 24.50
H2O 534 397
TEMP 350°F
FLUE GAS VOL 107,500 CF
LBS FLUE GAS 5,080
GRAINS/A.C.F 1.05
GRAINS/S.D.C.F 2.05
PARTICULATE 16.10 LBS
SO2 0.90 LBS
H2O 709 LBS
D.COLL.90%
OIL
COAL
NONE
D.COLL.93%
#6 OIL
COAL
TEMP. 350°F 350°F
FLUE GAS VOL 157,300CF 156,700CF
LBS FLUE GAS 7,640 7,860
GRAINS/A.C.F 0.02 1.91
GRAINS/S.D.C.F 0.04 3.19
PARTICULATE 0.47 42.90
SO2 18.70 24.50
H2O 534 397
TEMP 110 °F
GAS VOL 35 C.F.
H2O 0.20 LBS
PARTIC 0 LBS
H2S 0 LBS
RSH 0 LBS
RSR 0.10 LBS
RSSR 0.01 LBS
SO2 0 LBS
BARK = 900 LBS.
BARK
OIL/COAL
COMBINATION BOILER
#6 OIL
COAL
LBS 468 613
C 87.4% 72.0%
H 10.5% 5.5%
O 0% 10.5%
S 2.0% 2.0%
N 0% 0%
INERTS 0.1% 10.0%
BTU/LB 18,300 14,000
TEMP 120 °F
GAS VOL 50 C.F.
H2O 0.30 LBS
PARTIC 0 LBS
H2S 0.40 LBS
RSH 0.30 LBS
RSR 0.10 LBS
RSSR 0.01 LBS
SO2 0 LBS
SCRUBBER
C 29.3 %
H 8.1 %
O 60.9 %
S 0.1 %
N 0 %
INERTS 1.6 %
BTU/LB 4500
BLACK LIQUOR EVAPORATORS
WATER = 8 LBS.
SOLIDS = 127 LBS.
SOAP
WATER 12,903 LBS.
NA 17.70 %
C 39.20 %
H 3.93 %
O 33.15 %
S 4.15 %
N 0 %
INERTS 1.87 %
BTU/LB 6,600
TEMP 150 °F.
GAS FLOW, C.F. 12,200
GAS FLOW, LBS. 900
H2O 200 LBS
PARTIC 0 LBS
H2S 0.02 LBS
RSH 0.10 LBS
RSR 0.05 LBS
RSSR 0.16 LBS
SO2 0 LBS
TEMP 325 °F
GAS VOL 409,900 C.F.
H2O 4,230 LBS
PARTIC 106 LBS
H2S 1.50 LBS
RSH 0.10 LBS
RSR 0.01 LBS
RSSR 0.01 LBS
SO2 4.40 LBS
GRAINS/ACF 1.80
GRAINS/SDCF 4.00
GAS FLOW LBS 20,200
WATER = 2,729 LBS
SOLIDS = 2,729 LBS
CONCENTRATED BLACK LIQUOR OXIDATION SYSTEM 90% EFF
SOAP
WATER 2,529 LBS
SOLIDS = 2,829 LBS
AIR = 13,645 C.F. (1000 LBS)
190 F
44,500 C.F.
690 LBS
1.00 LBS
0.02 LBS
0.01 LBS
0.01 LBS
0 LBS
0 LBS
/ACF 0.20
/SDCF 0.30
OW LBS 2,320
NA 17.15 %
C 38.70 %
H 3.67 %
O 36.10 %
S 2.53 %
N 0 %
INERTS 1.85 %
BTU/LB 6,470
TEMP 200 °F
GAS VOL 45,000 C.F.
H2O 670 LBS
PARTIC 4.00 LBS
H2S 0.04 LBS
RSH 0.04 LBS
RSR 0.02 LBS
RSSR 0 LBS
SO2 0 LBS
GRAINS/ACF 0.60
GRAINS/SDCF 1.30
GAS FLOW LBS 2,320
LEGEND
PROCESS EQUIPMENT
CONTROL EQUIPMENT
AIR EMISSIONS

NOTES

1. ALL FLOWS REPRESENT THOSE FOR ONE AIR DRY TON OF UNBLEACHED PULP.
2. THE EMISSIONS FROM BURNING BARK AND OIL/COAL ARE BASED ON SUPPLYING THE TOTAL STEAM-ELECTRICAL REQUIREMENTS FOR PULP AND PAPER PRODUCTION (SEE "POWER PLANT ENERGY BALANCE"). OF THESE EMISSIONS, APPROXIMATELY 100% CAN BE ATTRIBUTED TO PAPER PRODUCTION AND 0% TO PULP PRODUCTION.
3. THE BARK FLOW IS BASED ON DEBARKING 100% OF THE WOOD SUPPLY. FOR MILLS RECEIVING WOOD CHIPS, THE BARK FLOW WILL BE REDUCED IN PROPORTION TO THE AMOUNT OF CHIPS BEING RECEIVED. THEREFORE, TO SATISFY THE STEAM REQUIREMENTS ON THE "POWER PLANT ENERGY BALANCE", THE OIL AND COAL FLOWS MUST BE INCREASED ACCORDINGLY.
4. IN IDENTIFYING THE AIR EMISSIONS THE METHYL FRACTION HAS BEEN ABBREVIATED TO "R". THEREFORE:
   RSH = METHYL MERCAPTAN
   RSR = DIMETHYL SULFIDE
   RSSR = DIMETHYL DISULFIDE
5. ABBREVIATIONS:
   C. F. AND A. C. F. = ACTUAL CUBIC FEET AT FLOW CONDITIONS
   S. D. C. F. = STANDARD DRY CUBIC FEET AT 70° F. AND 14.7 P.S.I.A.
   EFF. = ANNUAL OPERATING EFFICIENCY AT THE CONTROL EQUIPMENT
6. ALL VALUES ARE ANNUAL AVERAGES.
7. ALL FUEL ANALYSES ARE ON AN "AS FIRED" BASIS.

KRAFT FLOW DIAGRAM NO.4

CONTINUOUS DIGESTER, PAPER STOCK, HAVING CONCENTRATED BLACK LIQUOR OXIDATION, DIRECT CONTACT EVAPORATOR, LIME KILN HIGH EFFICIENCY COLLECTION AND NO BLEACHING

EXHIBIT NO.

SYSTEMS ANALYSIS STUDY OF EMISSIONS CONTROL IN THE WOOD PULPING INDUSTRY

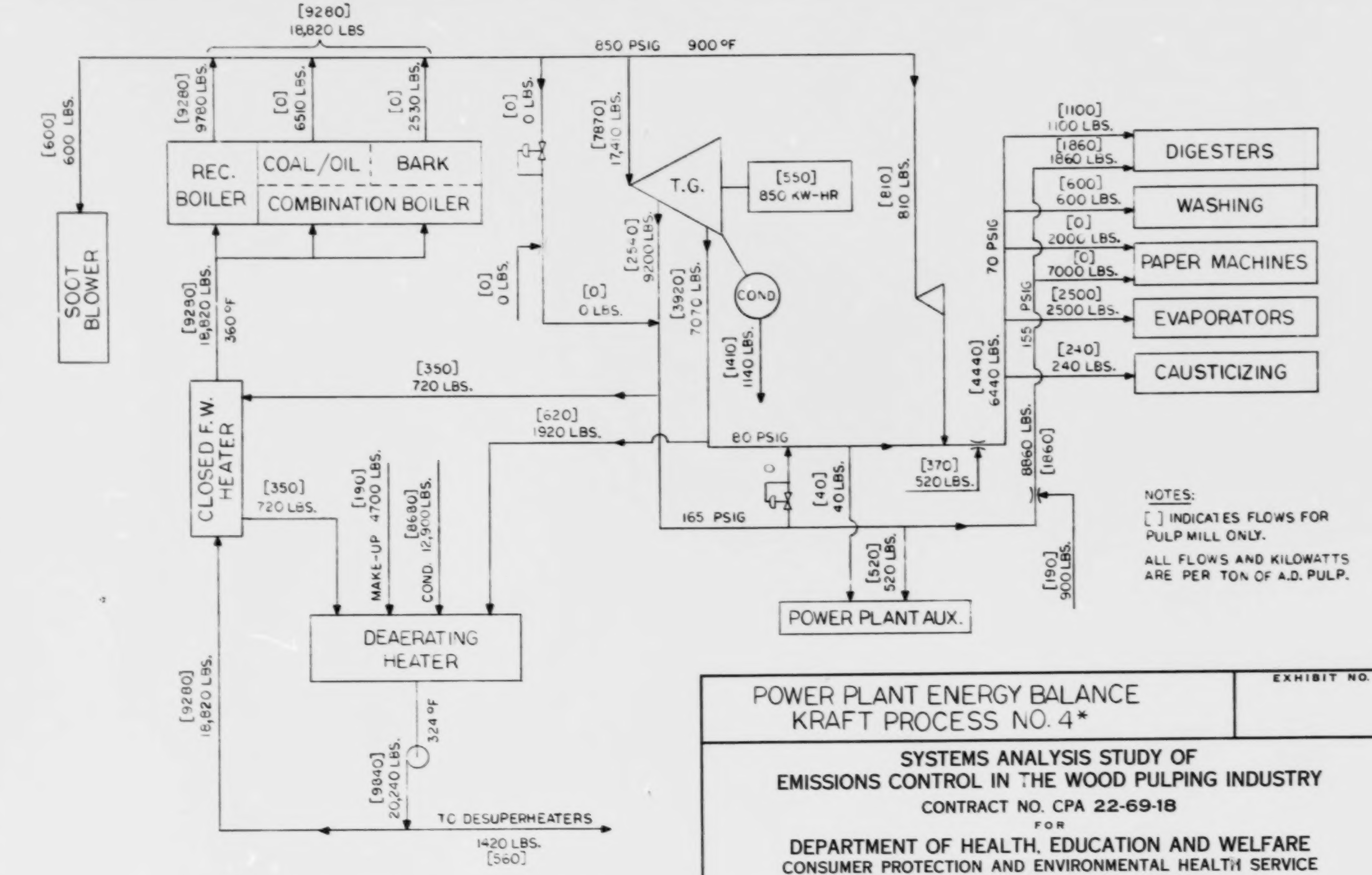
CONTRACT NO. CPA 22-69-18

FOR

DEPARTMENT OF HEALTH, EDUCATION AND WELFARE
CONSUMER PROTECTION AND ENVIRONMENTAL HEALTH SERVICE
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

ENVIRONMENTAL ENGINEERING, INC.
GAINESVILLE, FLORIDA

J. E. SIRRINE COMPANY, ENGINEERS
GREENVILLE, S. C.


[9280]
18,820 LBS
850 PSIG 900 °F
[600]
600 LBS.
SOOT BLOWER
[9280]
9780 LBS.
[0]
6510 LBS.
[0]
2530 LBS.
REC. BOILER
COAL/OIL
BARK
COMBINATION BOILER
[9280]
18,820 LBS.
360 °F
[0]
0 LBS.
[0]
0 LBS.
[0]
0 LBS.
[7870]
17,410 LBS.
[25+0]
9200 LBS.
T.G.
[550]
850 KW-HR
[3920]
7070 LBS.
COND
[1410]
1140 LBS.
[810]
810 LBS.
70 PSIG
155 PSIG
[1100]
1100 LBS.
DIGESTERS
[1860]
1860 LBS.
[600]
600 LBS.
WASHING
[0]
2000 LBS.
[0]
7000 LBS.
PAPER MACHINES
[2500]
2500 LBS.
EVAPORATORS
[240]
240 LBS.
CAUSTICIZING
[4440]
6440 LBS.
[350]
720 LBS.
[620]
1920 LBS.
80 PSIG
[370]
520 LBS.
8860 LBS.
[1860]
0
[40]
40 LBS.
165 PSIG
[520]
520 LBS.
POWER PLANT AUX.
[190]
900 LBS.
CLOSED F.W. HEATER
[350]
720 LBS.
[190]
MAKE-UP 4700 LBS.
[8680]
COND. 12,900 LBS.
DEAERATING HEATER
[9280]
18,820 LBS.
324 °F
[9840]
20,240 LBS.
TO DESUPERHEATERS
1420 LBS.
[560]

NOTES:
[ ] INDICATES FLOWS FOR PULP MILL ONLY.
ALL FLOWS AND KILOWATTS ARE PER TON OF A.D. PULP.

* CONTINUOUS DIGESTERS, CONCENTRATED BLACK LIQUOR OXIDATION, DIRECT CONTACT EVPORATORS, NO BLEACHING, HIGH EFFICIENCY LIME KILN COLLECTION.

POWER PLANT ENERGY BALANCE KRAFT PROCESS NO. 4*

EXHIBIT NO.

SYSTEMS ANALYSIS STUDY OF EMISSIONS CONTROL IN THE WOOD PULPING INDUSTRY

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FOR
DEPARTMENT OF HEALTH, EDUCATION AND WELFARE
CONSUMER PROTECTION AND ENVIRONMENTAL HEALTH SERVICE
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

ENVIRONMENTAL ENGINEERING, INC.
GAINESVILLE, FLORIDA

J. E. SIRRINE COMPANY, ENGINEERS
GREENVILLE, S. C.

KRAFT DIAGRAM NO. 5

TYPICAL LOCATION

Southern United States.

TYPICAL AGE OF EQUIPMENT

Over 15 years.

GENERAL

This flow diagram is based on production of top liner stock and should be used with Flow Diagram No. 1, base stock, to determine air emissions from a linerboard mill. Flow Diagram No. 5 depicts a relatively old mill, using batch digesters, and with no provisions for odor abatement. There are perhaps 30 to 40 mills in the U. S. which are illustrated by this flow diagram.

EMISSIONS

The following emissions have been selected as representative, based on the range of emissions presented in Chapter 4.

KRAFT DIAGRAM NO. 5
(Continued)

EMISSIONS - Continued

POUNDS PER AIR DRY TON OF UNBLEACHED PULP

LOCATION	H_2S	RSH, RSR, RSSR	SO_2	Particulate
Accumulator	0.12	3.35	0	0
Washers and Screens	0.02	0.17	0	0
M. E. Evaporators	0.50	0.31	0	0
R. Boilers and D. C. Evaporators	14.6	1.50	4.8	91.8
After Precipitators	14.6	1.50	4.8	11.0
Smelt Dissolving Tank	0.04	0.05	0	4.0
Slaker	0	0	0	Unknown
Lime Kiln	1.0	0.6	0.1	51
Lime Kiln Scrubber	0.40	0.23	Trace	10.0
Turpentine Condenser	0.01	0.50	0	0

ASSUMPTIONS

The following assumptions have been made in developing the flow diagram:

A. PULP MILL

1. Pulp Yield = 47 Percent.
2. Cooking liquor charge per air dry ton of brown stock equals 6,335 pounds of which 900 pounds are chemical solids. Sulfidity equals 25 percent.
3. Black liquor solids from first stage washer equals 15 percent and from evaporators equals 50 percent solids.

KRAFT DIAGRAM NO. 5
(Continued)

ASSUMPTIONS - Continued

B. RECOVERY

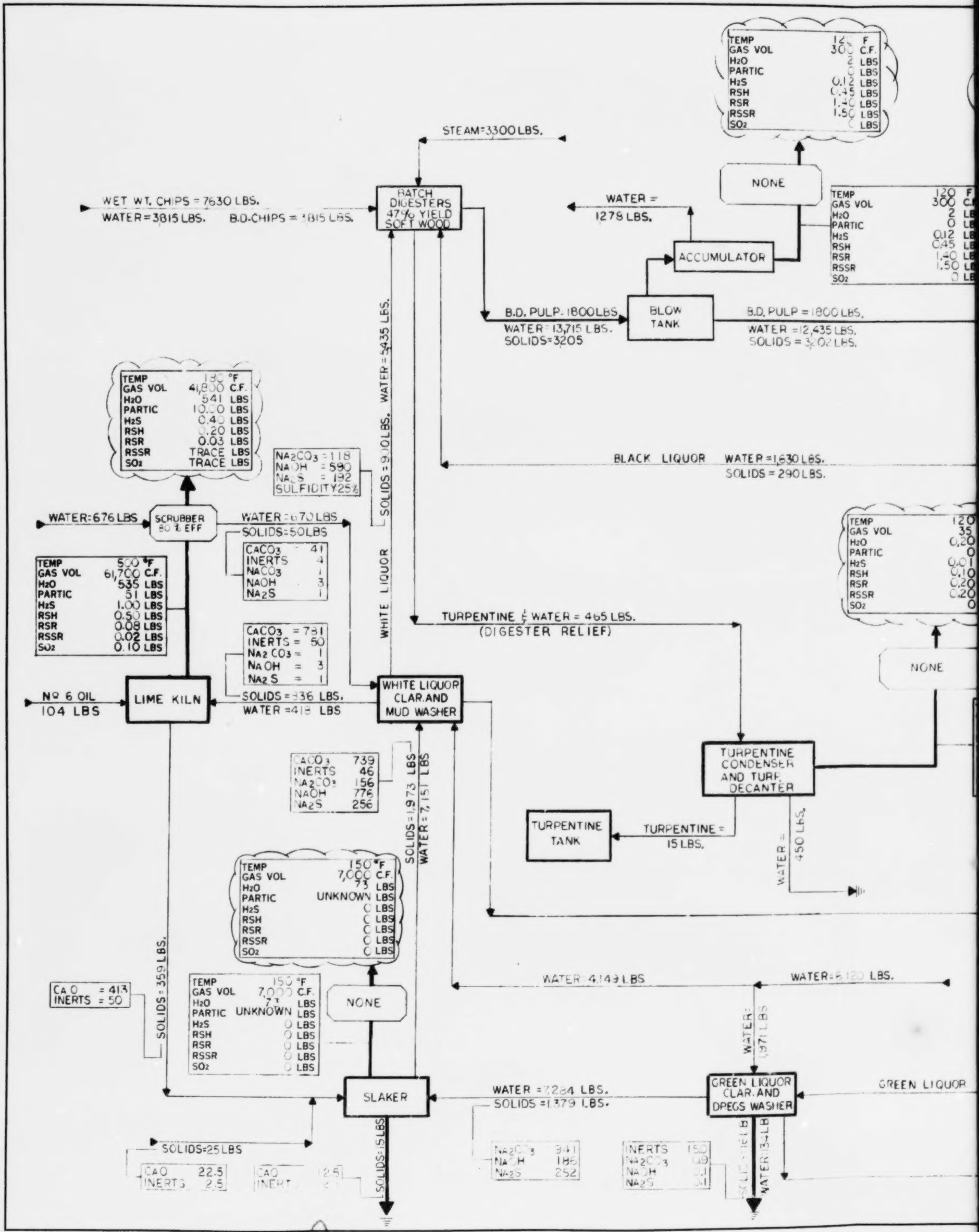
1. Unit operating at 15 percent excess air at economizer outlet.
2. Particulate matter of 8 grains per SDCF leaving the economizer.
3. Efficiency of 50 percent on particulate removal in the direct contact evaporators.
4. Steam shatter jets utilized on smelt spouts.
5. Design efficiency of 95 percent for precipitator with an annual average operating efficiency of 88 percent.

C. CAUSTICIZING

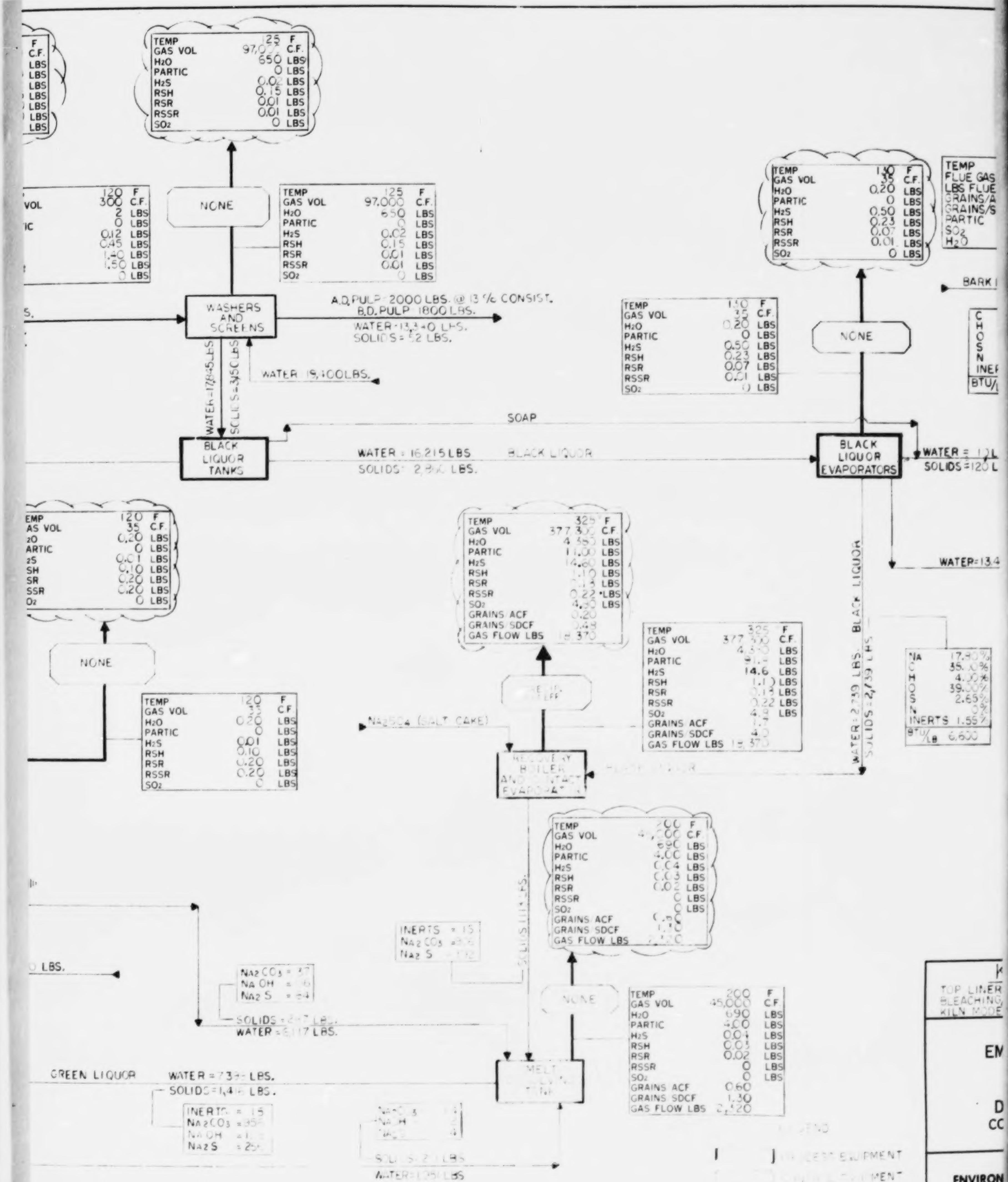
1. Lime kiln SO_2 emission has been assumed at 0.1 pounds per air dry ton of pulp.
2. Lime kiln scrubber efficiency assumed at 80 percent on lime and 60 percent efficiency on soda.

D. POWER PLANT

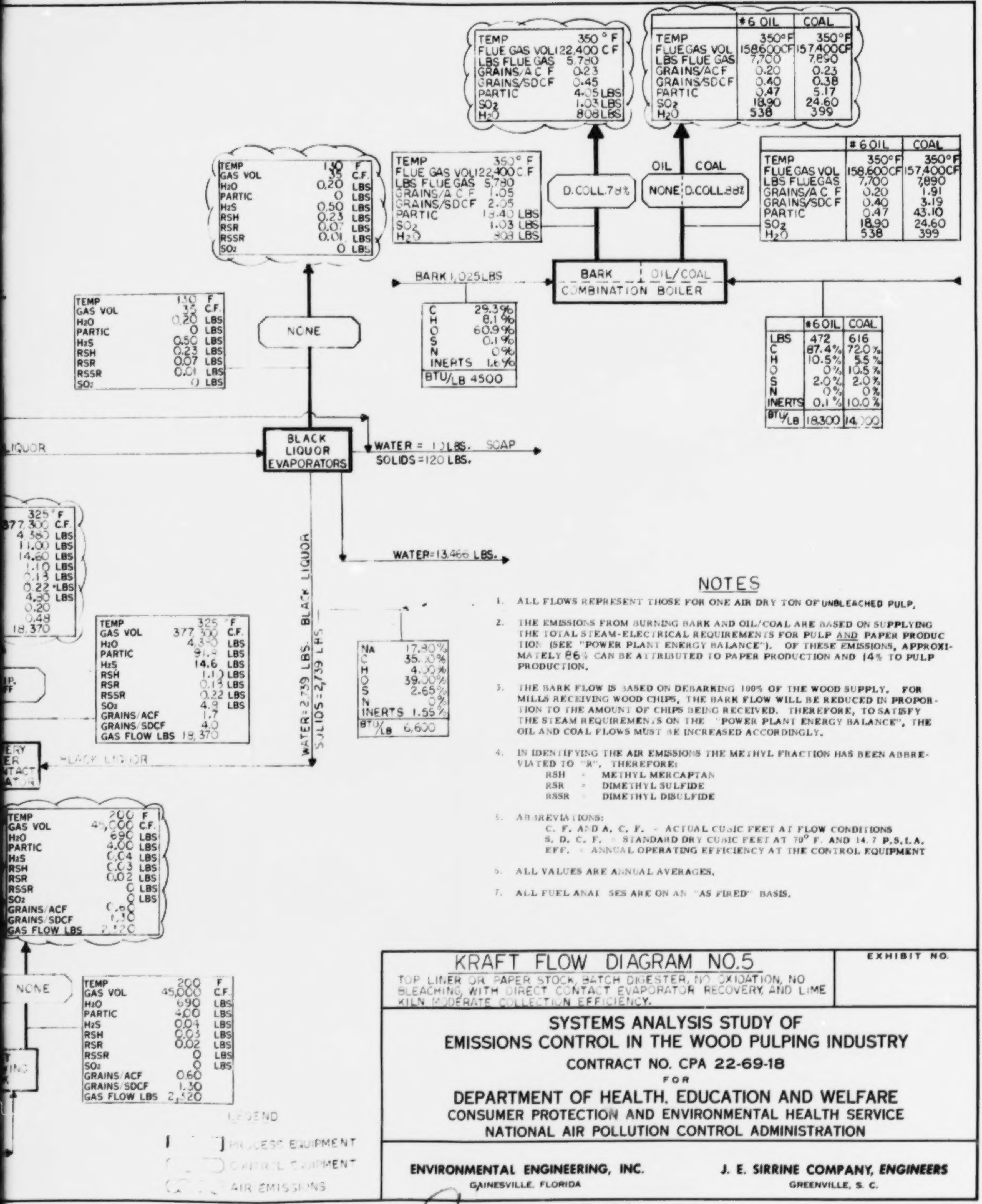
1. Excess air: 30 percent for bark, 10 percent for oil and 20 percent for coal.
2. Bark burning equipment is a low set spreader stoker.
3. Reinjection from dust collector: 50 percent for bark and 0 percent for coal.
4. Design efficiencies for dust collector: 80 percent on bark and 90 percent on coal with annual operating efficiencies of 78 percent and 88 percent, respectively.
5. Bark is burned 24 hours per day at a controlled rate.
6. Unburned combustible in the refuse leaving the dust collector is 40 percent.
7. Coal firing based on pulverized coal.


STEAM=3300 LBS.
WET WT. CHIPS = 7630 LBS.
WATER=3815 LBS.
B.D.CHIPS = 3815 LBS.
BATCH DIGESTERS 47% YIELD SOFT WOOD
WATER = 1278 LBS.
ACCUMULATOR
NONE
TEMP 120 F
GAS VOL 300 C.F.
H2O 2 LBS
PARTIC 0 LBS
H2S 0.12 LBS
RSH 0.45 LBS
RSR 1.40 LBS
RSSR 1.50 LBS
SO2 0 LBS
TEMP 120 F
GAS VOL 300 C.
H2O 2 LB
PARTIC 0 LB
H2S 0.12 LB
RSH 0.45 LB
RSR 1.40 LB
RSSR 1.50 LB
SO2 0 LB
B.D. PULP=1800 LBS.
WATER=13,715 LBS.
SOLIDS=3205
BLOW TANK
B.D. PULP = 1800 LBS.
WATER = 12,435 LBS.
SOLIDS = 3202 LBS.
TEMP 180 °F
GAS VOL 41,800 C.F.
H2O 541 LBS
PARTIC 10.00 LBS
H2S 0.40 LBS
RSH 0.20 LBS
RSR 0.03 LBS
RSSR TRACE LBS
SO2 TRACE LBS
NA2CO3 = 118
NAOH = 590
NA2S = 192
SULFIDITY 25%
SOLIDS = 900 LBS.
WATER = 5435 LBS.
WHITE LIQUOR
BLACK LIQUOR
WATER = 1,630 LBS.
SOLIDS = 290 LBS.
WATER: 676 LBS
SCRUBBER 80% EFF
WATER = 670 LBS
SOLIDS = 50 LBS
CACO3 41
INERTS 4
NACO3 1
NAOH 3
NA2S 1
TEMP 500 °F
GAS VOL 61,700 C.F.
H2O 535 LBS
PARTIC 51 LBS
H2S 1.00 LBS
RSH 0.50 LBS
RSR 0.08 LBS
RSSR 0.02 LBS
SO2 0.10 LBS
CACO3 = 781
INERTS = 50
NA2 CO3 = 1
NA OH = 3
NA2 S = 1
TEMP 120
GAS VOL 35
H2O 0.20
PARTIC 0
H2S 0.01
RSH 0.10
RSR 0.20
RSSR 0.20
SO2 0
TURPENTINE & WATER = 465 LBS.
(DIGESTER RELIEF)
NONE
Nº 6 OIL
104 LBS
LIME KILN
SOLIDS = 336 LBS.
WATER = 418 LBS
WHITE LIQUOR CLAR. AND MUD WASHER
CACO3 739
INERTS 46
NA2CO3 156
NAOH 776
NA2S 256
SOLIDS = 1,973 LBS
WATER = 7,151 LBS
TURPENTINE CONDENSER AND TURP. DECANTER
TURPENTINE TANK
TURPENTINE = 15 LBS.
WATER = 450 LBS.
TEMP 150 °F
GAS VOL 7,000 C.F.
H2O 73 LBS
PARTIC UNKNOWN LBS
H2S 0 LBS
RSH 0 LBS
RSR 0 LBS
RSSR 0 LBS
SO2 0 LBS
CA O = 413
INERTS = 50
SOLIDS = 359 LBS.
TEMP 150 °F
GAS VOL 7,000 C.F.
H2O 73 LBS
PARTIC UNKNOWN LBS
H2S 0 LBS
RSH 0 LBS
RSR 0 LBS
RSSR 0 LBS
SO2 0 LBS
NONE
WATER = 4,149 LBS
WATER = 6,120 LBS.
WATER: 1,971 LBS
SLAKER
WATER = 7,284 LBS.
SOLIDS = 1,379 LBS.
GREEN LIQUOR CLAR. AND DREGS WASHER
GREEN LIQUOR
SOLIDS = 25 LBS
CAO 22.5
INERTS 2.5
CAO 12.5
INERTS 2.5
SOLIDS: 15 LBS
NA2CO3 941
NAOH 186
NA2S 252
INERTS 15.0
NA2CO3 0.9
NAOH 0.1
NA2S 0.1
SOLIDS: 16 LB
WATER: 134 LB

A


TEMP 125 F
GAS VOL 97,000 C.F.
H2O 650 LBS
PARTIC 0 LBS
H2S 0.02 LBS
RSH 0.15 LBS
RSR 0.01 LBS
RSSR 0.01 LBS
SO2 0 LBS
NONE
WASHERS AND SCREENS
A.D. PULP 2000 LBS. @ 13 % CONSIST.
B.D. PULP 1800 LBS.
WATER 13,340 LBS.
SOLIDS = 52 LBS.
WATER 19,400 LBS.
WATER = 17,845 LBS
SOLIDS = 350 LBS
SOAP
BLACK LIQUOR TANKS
WATER = 16,215 LBS
BLACK LIQUOR
SOLIDS = 2,850 LBS.
TEMP 130 F
GAS VOL 35 C.F.
H2O 0.20 LBS
PARTIC 0 LBS
H2S 0.50 LBS
RSH 0.23 LBS
RSR 0.07 LBS
RSSR 0.01 LBS
SO2 0 LBS
NONE
BLACK LIQUOR EVAPORATORS
WATER = 13,4
BLACK LIQUOR
WATER = 2,739 LBS.
SOLIDS = 2,739 LBS.
BARK
TEMP
FLUE GAS
LBS FLUE
GRAINS/A
GRAINS/S
PARTIC
SO2
H2O
TEMP 120 F
GAS VOL 35 C.F.
H2O 0.20 LBS
PARTIC 0 LBS
H2S 0.01 LBS
RSH 0.10 LBS
RSR 0.20 LBS
RSSR 0.20 LBS
SO2 0 LBS
NONE
TEMP 325 F
GAS VOL 377,300 C.F.
H2O 4,380 LBS
PARTIC 11.00 LBS
H2S 14.60 LBS
RSH 1.10 LBS
RSR 0.13 LBS
RSSR 0.22 LBS
SO2 4.80 LBS
GRAINS ACF 0.20
GRAINS SDCF 0.48
GAS FLOW LBS 19,370
PARTIC 91.6 LBS
GRAINS ACF 1.7
GRAINS SDCF 4.0
Na2SO4 (SALT CAKE)
RECOVERY BOILER AND CONTACT EVAPORATOR
NA 17.40%
C 35.00%
H 4.00%
O 39.00%
S 2.65%
N 0%
INERTS 1.55%
BTU/LB 6,600
TEMP 200 F
GAS VOL 45,000 C.F.
H2O 690 LBS
PARTIC 4.00 LBS
H2S 0.04 LBS
RSH 0.03 LBS
RSR 0.02 LBS
RSSR 0 LBS
SO2 0 LBS
GRAINS ACF 0.60
GRAINS SDCF 1.30
GAS FLOW LBS 2,420
NONE
SOLIDS LBS.
INERTS = 15
Na2CO3 = 808
Na2S = 172
Na2CO3 = 37
NaOH = 16
Na2S = 64
SOLIDS = 247 LBS.
WATER = 6,117 LBS.
MELT DISSOLVING TANK
GREEN LIQUOR WATER = 7,33 LBS.
SOLIDS = 1,41 LBS.
INERTS = 15
Na2CO3 = 858
NaOH = 1
Na2S = 25
SOLIDS = 2 LBS
WATER = 1,351 LBS
LEGEND
PROCESS EQUIPMENT
CONTROL EQUIPMENT
AIR EMISSIONS
TOP LINER
BLEACHING
KILN MODE
EM
D
CC
ENVIRON

B



TEMP	130	F
GAS VOL	35	C.F.
H2O	0.20	LBS
PARTIC	0	LBS
H2S	0.50	LBS
RSH	0.23	LBS
RSR	0.07	LBS
RSSR	0.01	LBS
SO2	0	LBS

TEMP	130	F
GAS VOL	35	C.F.
H2O	0.20	LBS
PARTIC	0	LBS
H2S	0.50	LBS
RSH	0.23	LBS
RSR	0.07	LBS
RSSR	0.01	LBS
SO2	0	LBS

TEMP	350 °F
FLUE GAS VOL	122,400 CF
LBS FLUE GAS	5,780
GRAINS/A C F	0.23
GRAINS/SDCF	0.45
PARTIC	4.05 LBS
SO2	1.03 LBS
H2O	808 LBS

TEMP	350 °F
FLUE GAS VOL	122,400 C.F
LBS FLUE GAS	5,780
GRAINS/A C F	1.05
GRAINS/SDCF	2.05
PARTIC	18.40 LBS
SO2	1.03 LBS
H2O	808 LBS

	#6 OIL	COAL
TEMP	350°F	350°F
FLUE GAS VOL	158600CF	157,400CF
LBS FLUE GAS	7,700	7,890
GRAINS/ACF	0.20	0.23
GRAINS/SDCF	0.40	0.38
PARTIC	0.47	5.17
SO2	18.90	24.60
H2O	538	399

	#6 OIL	COAL
TEMP	350°F	350°F
FLUE GAS VOL	158,600CF	157,400CF
LBS FLUE GAS	7,700	7,890
GRAINS/A C F	0.20	1.91
GRAINS/SDC F	0.40	3.19
PARTIC	0.47	43.10
SO2	18.90	24.60
H2O	538	399

C	29.3%
H	8.1%
O	60.9%
S	0.1%
N	0%
INERTS	1.6%
BTU/LB	4500

	#6 OIL	COAL
LBS	472	616
C	87.4%	72.0%
H	10.5%	5.5%
O	0%	10.5%
S	2.0%	2.0%
N	0%	0%
INERTS	0.1%	10.0%
BTU/LB	18,300	14,000

NA	17.80%
C	35.00%
H	4.00%
O	39.00%
S	2.65%
N	0%
INERTS	1.55%
BTU/LB	6,600

	325	°F
	377,300	C.F.
	4,380	LBS
	11.00	LBS
	14.60	LBS
	1.10	LBS
	0.13	LBS
	0.22	LBS
	4.30	LBS
	0.20	
	0.48	
	18,370	

TEMP	325	°F
GAS VOL	377,300	C.F.
H2O	4,380	LBS
PARTIC	91.9	LBS
H2S	14.6	LBS
RSH	1.10	LBS
RSR	0.13	LBS
RSSR	0.22	LBS
SO2	4.9	LBS
GRAINS/ACF	1.7	
GRAINS/SDCF	4.0	
GAS FLOW LBS	18,370	

TEMP	200	F
GAS VOL	45,000	C.F.
H2O	690	LBS
PARTIC	4.00	LBS
H2S	0.04	LBS
RSH	0.03	LBS
RSR	0.02	LBS
RSSR	0	LBS
SO2	0	LBS
GRAINS/ACF	0.60	
GRAINS/SDCF	1.30	
GAS FLOW LBS	2,320	

TEMP	200	F
GAS VOL	45,000	C.F.
H2O	690	LBS
PARTIC	4.00	LBS
H2S	0.04	LBS
RSH	0.03	LBS
RSR	0.02	LBS
RSSR	0	LBS
SO2	0	LBS
GRAINS/ACF	0.60	
GRAINS/SDCF	1.30	
GAS FLOW LBS	2,320	

NOTES

1. ALL FLOWS REPRESENT THOSE FOR ONE AIR DRY TON OF UNBLEACHED PULP.
2. THE EMISSIONS FROM BURNING BARK AND OIL/COAL ARE BASED ON SUPPLYING THE TOTAL STEAM-ELECTRICAL REQUIREMENTS FOR PULP AND PAPER PRODUCTION (SEE "POWER PLANT ENERGY BALANCE"). OF THESE EMISSIONS, APPROXIMATELY 86% CAN BE ATTRIBUTED TO PAPER PRODUCTION AND 14% TO PULP PRODUCTION.
3. THE BARK FLOW IS BASED ON DEBARKING 100% OF THE WOOD SUPPLY. FOR MILLS RECEIVING WOOD CHIPS, THE BARK FLOW WILL BE REDUCED IN PROPORTION TO THE AMOUNT OF CHIPS BEING RECEIVED. THEREFORE, TO SATISFY THE STEAM REQUIREMENTS ON THE "POWER PLANT ENERGY BALANCE", THE OIL AND COAL FLOWS MUST BE INCREASED ACCORDINGLY.
4. IN IDENTIFYING THE AIR EMISSIONS THE METHYL FRACTION HAS BEEN ABBREVIATED TO "R". THEREFORE:
    RSH = METHYL MERCAPTAN
    RSR = DIMETHYL SULFIDE
    RSSR = DIMETHYL DISULFIDE
5. ABBREVIATIONS:
    C. F. AND A. C. F. = ACTUAL CUBIC FEET AT FLOW CONDITIONS
    S. D. C. F. = STANDARD DRY CUBIC FEET AT 70° F. AND 14.7 P.S.I.A.
    EFF. = ANNUAL OPERATING EFFICIENCY AT THE CONTROL EQUIPMENT
6. ALL VALUES ARE ANNUAL AVERAGES.
7. ALL FUEL ANALYSES ARE ON AN "AS FIRED" BASIS.

KRAFT FLOW DIAGRAM NO.5

TOP LINER OR PAPER STOCK, BATCH DIGESTER, NO OXIDATION, NO BLEACHING, WITH DIRECT CONTACT EVAPORATOR RECOVERY, AND LIME KILN MODERATE COLLECTION EFFICIENCY.

EXHIBIT NO.

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EMISSIONS CONTROL IN THE WOOD PULPING INDUSTRY
CONTRACT NO. CPA 22-69-18
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GAINESVILLE, FLORIDA

J. E. SIRRINE COMPANY, ENGINEERS
GREENVILLE, S. C.

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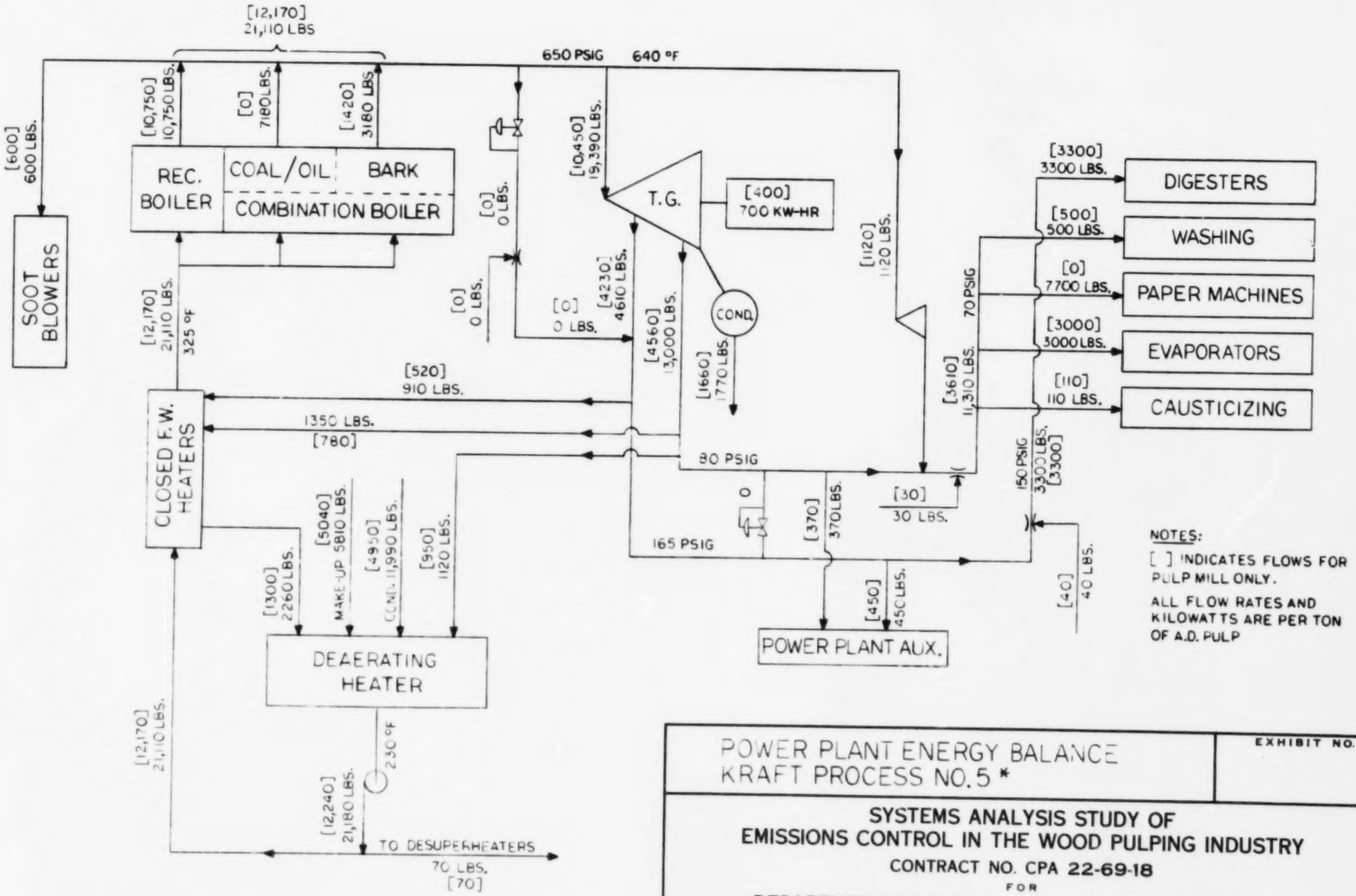

[12,170]
21,110 LBS
650 PSIG
640 °F
[600]
600 LBS.
SOOT BLOWERS
[10,750]
10,750 LBS.
[0]
7180 LBS.
[1420]
3180 LBS
REC. BOILER
COAL/OIL
BARK
COMBINATION BOILER
[10,450]
19,390 LBS.
T.G.
[400]
700 KW-HR
[0]
0 LBS.
[0]
0 LBS.
[0]
0 LBS.
[4230]
4610 LBS.
[4560]
13,000 LBS.
COND.
[1660]
1770 LBS.
[1120]
1120 LBS.
[12,170]
21,110 LBS.
325 °F
[520]
910 LBS.
1350 LBS.
[780]
CLOSED F.W. HEATERS
80 PSIG
165 PSIG
0
[1300]
2260 LBS.
[5040]
MAKE-UP 5810 LBS.
[4950]
COND. 11,990 LBS.
[950]
1120 LBS.
DEAERATING HEATER
230 °F
[12,170]
21,110 LBS.
[12,240]
21,180 LBS.
TO DESUPERHEATERS
70 LBS.
[70]
[370]
370 LBS.
[450]
450 LBS.
POWER PLANT AUX.
[30]
30 LBS.
[3610]
11,310 LBS.
70 PSIG
150 PSIG
3300 LBS.
[3300]
[40]
40 LBS.
[3300]
3300 LBS.
DIGESTERS
[500]
500 LBS.
WASHING
[0]
7700 LBS.
PAPER MACHINES
[3000]
3000 LBS.
EVAPORATORS
[110]
110 LBS.
CAUSTICIZING

NOTES:

[ ] INDICATES FLOWS FOR PULP MILL ONLY.

ALL FLOW RATES AND KILOWATTS ARE PER TON OF A.D. PULP

* TOP LINER OR PAPER STOCK, BATCH DIGESTERS, NO OXIDATION, DIRECT CONTACT EVAPORATOR RECOVERY, NO BLEACHING, AND LIME KILN MODERATE COLLECTION EFFICIENCY.

POWER PLANT ENERGY BALANCE
KRAFT PROCESS NO. 5 *

EXHIBIT NO.

SYSTEMS ANALYSIS STUDY OF
EMISSIONS CONTROL IN THE WOOD PULPING INDUSTRY
CONTRACT NO. CPA 22-69-18
FOR
DEPARTMENT OF HEALTH, EDUCATION AND WELFARE
CONSUMER PROTECTION AND ENVIRONMENTAL HEALTH SERVICE
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

ENVIRONMENTAL ENGINEERING, INC.
GAINESVILLE, FLORIDA

J. E. SIRRINE COMPANY, ENGINEERS
GREENVILLE, S. C.

KRAFT DIAGRAM NO. 6

TYPICAL LOCATION

Southern United States.

TYPICAL AGE OF EQUIPMENT

Five to ten years.

GENERAL

This flow diagram is based on cooking in an early design continuous type digester with no washing. A bleach plant is included. The flow diagram depicts a mill with minor provisions for odor abatement. There are perhaps 10 to 20 mills in the U. S. that are illustrated by this flow diagram.

EMISSIONS

The following emissions have been selected as representative, based on the range of emissions presented in Chapter 4.

POUNDS PER AIR DRY TON OF UNBLEACHED PULP

LOCATION	H_2S	RSH, RSR, RSSR	SO_2	Particulate
Washers and Screens	0.02	0.23	0	0
M. E. Evaporators	0.50	0.28	0	0
R. Boiler and D. C. Evaporator	17.5	1.95	5.0	118
After Precipitators	17.5	1.95	5.0	5.90
Smelt Dissolving Tank	0.02	0.02	0	1.0
Slaker	0	0	0	Unknown
Lime Kiln	1.0	0.5	Trace	31.5
Lime Kiln Scrubber	0.40	0.20	Trace	6.3
Turpentine Condenser	0.01	0.30	0	0

KRAFT DIAGRAM NO. 6
(Continued)

ASSUMPTIONS

The following assumptions have been made in developing the flow diagram:

A. PULP MILL

1. Pulp Yield = 45 Percent.
2. Cooking liquor charge per air dry ton of brown stock equals 8,106 pounds of which 1,150 pounds are chemical solids. Sulfidity equals 27 percent.
3. Black liquor solids from first stage washer equals 16 percent and from evaporators equals 50 percent solids.

B. BLEACH PLANT

1. Assumed a five stage Bleach Plant with a chlorine emission of one pound per air dry ton of pulp.

C. RECOVERY

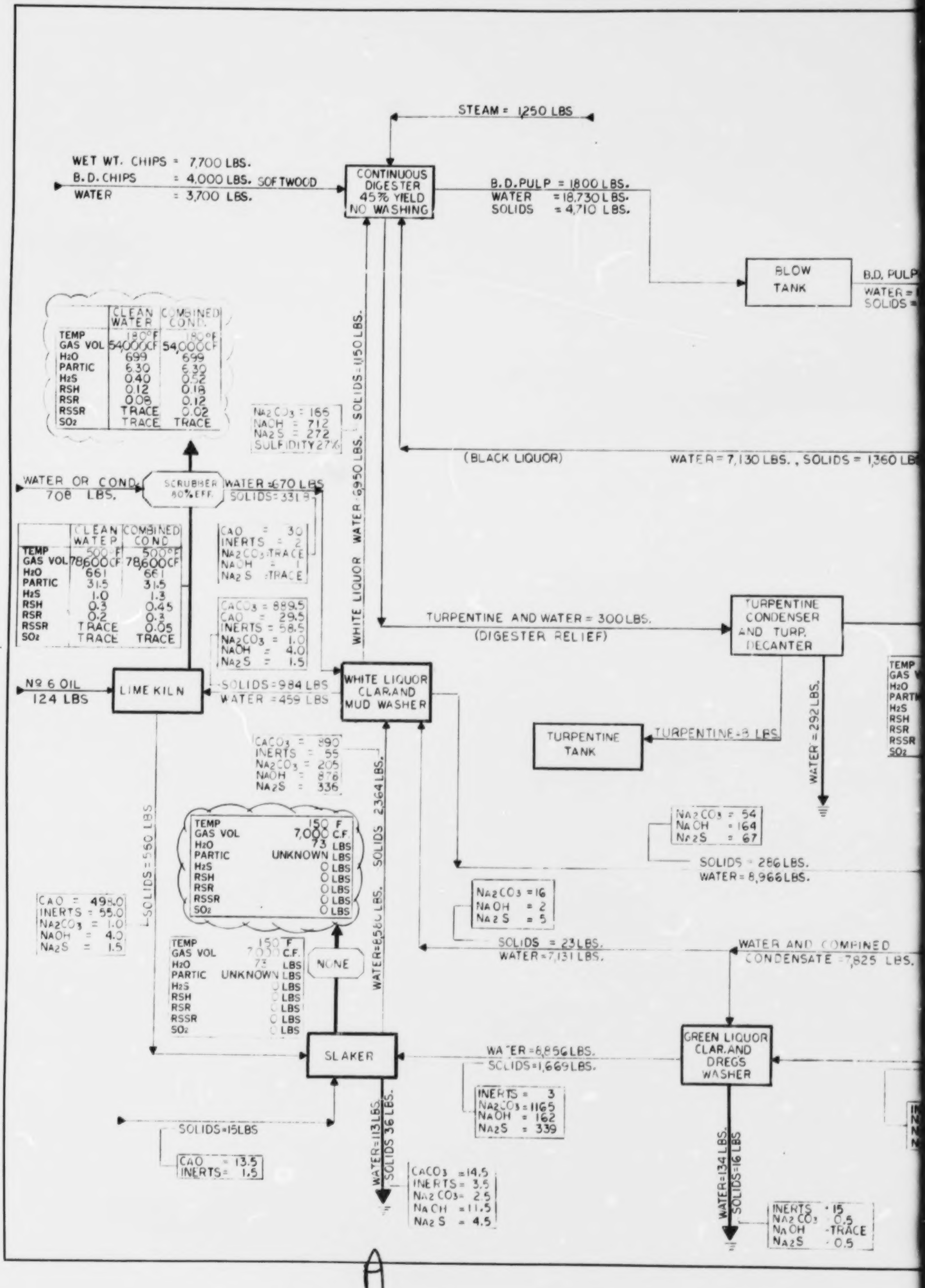
1. Unit operating at 15 percent excess air at economizer outlet.
2. Particulate matter of 8 grains per SDCF leaving the economizer.
3. Efficiency of 50 percent on particulate removal in the direct contact evaporators.
4. Steam shatter jets utilized on smelt spouts.
5. Design efficiency of 98 percent for precipitator with an annual average operating efficiency of 95 percent.

D. CAUSTICIZING

1. Lime kiln scrubber efficiency assumed at 80 percent on lime and 60 percent efficiency on soda.

E. POWER PLANT

1. Excess air: 30 percent for bark, 10 percent for oil and 20 percent for coal.
2. Bark burning equipment is a low set spreader stoker.
3. Reinjection from dust collector: 50 percent for bark and 0 percent for coal.
4. Design efficiencies for dust collector: 92 percent on bark and 95 percent on coal with annual operating efficiencies of 90 percent and 93 percent, respectively.
5. Bark is burned 24 hours per day at a controlled rate.
6. Unburned combustible in refuse leaving the dust collector is 40 percent.
7. Coal firing based on pulverized coal.


STEAM = 1250 LBS
WET WT. CHIPS = 7,700 LBS.
B.D. CHIPS = 4,000 LBS. SOFTWOOD
WATER = 3,700 LBS.
CONTINUOUS DIGESTER 45% YIELD NO WASHING
B.D. PULP = 1800 LBS.
WATER = 18,730 LBS.
SOLIDS = 4,710 LBS.
BLOW TANK
B.D. PULP
WATER =
SOLIDS =
WHITE LIQUOR WATER 6950 LBS. SOLIDS = 1150 LBS.
(BLACK LIQUOR)
WATER = 7,130 LBS., SOLIDS = 1,360 LB
WATER OR COND. 708 LBS.
SCRUBBER 80% EFF.
WATER = 670 LBS
SOLIDS = 33 LB
Nº 6 OIL 124 LBS
LIME KILN
SOLIDS = 984 LBS
WATER = 459 LBS
WHITE LIQUOR CLAR. AND MUD WASHER
TURPENTINE AND WATER = 300 LBS.
(DIGESTER RELIEF)
TURPENTINE CONDENSER AND TURP. DECANTER
TURPENTINE TANK
TURPENTINE = 8 LBS
WATER = 292 LBS.
SOLIDS = 560 LBS
SOLIDS 2364 LBS.
WATER = 6,580 LBS.
SOLIDS = 286 LBS.
WATER = 8,966 LBS.
SOLIDS = 23 LBS.
WATER = 7,131 LBS.
WATER AND COMBINED CONDENSATE = 7,825 LBS.
NONE
SLAKER
WATER = 6,856 LBS.
SOLIDS = 1,669 LBS.
GREEN LIQUOR CLAR. AND DREGS WASHER
SOLIDS = 15 LBS
WATER = 113 LBS.
SOLIDS 36 LBS.
WATER = 134 LBS.
SOLIDS = 16 LBS
TEMP
GAS
H2O
PARTI
H2S
RSH
RSR
RSSR
SO2
IN
NA
NA
NA

	CLEAN WATER	COMBINED COND.
TEMP	180°F	180°F
GAS VOL	54,000CF	54,000CF
H_2O	699	699
PARTIC	6.30	6.30
H_2S	0.40	0.52
RSH	0.12	0.18
RSR	0.08	0.12
RSSR	TRACE	0.02
SO_2	TRACE	TRACE

	CLEAN WATER	COMBINED COND.
TEMP	500°F	500°F
GAS VOL	78,600CF	78,600CF
H_2O	661	661
PARTIC	31.5	31.5
H_2S	1.0	1.3
RSH	0.3	0.45
RSR	0.2	0.3
RSSR	TRACE	0.05
SO_2	TRACE	TRACE

Na_2CO_3 = 166
NAOH = 712
Na_2S = 272
SULFIDITY 27%

CAO = 30
INERTS = 2
Na_2CO_3 = TRACE
NAOH = 1
Na_2S = TRACE

$CaCO_3$ = 889.5
CAO = 29.5
INERTS = 58.5
Na_2CO_3 = 1.0
NAOH = 4.0
Na_2S = 1.5

$CaCO_3$ = 890
INERTS = 55
Na_2CO_3 = 205
NAOH = 876
Na_2S = 336

TEMP	150	F
GAS VOL	7,000	C.F.
H_2O	73	LBS
PARTIC	UNKNOWN	LBS
H_2S	0	LBS
RSH	0	LBS
RSR	0	LBS
RSSR	0	LBS
SO_2	0	LBS

TEMP	150	F
GAS VOL	7,000	C.F.
H_2O	73	LBS
PARTIC	UNKNOWN	LBS
H_2S	0	LBS
RSH	0	LBS
RSR	0	LBS
RSSR	0	LBS
SO_2	0	LBS

CAO = 498.0
INERTS = 55.0
Na_2CO_3 = 1.0
NAOH = 4.0
Na_2S = 1.5

Na_2CO_3 = 54
NAOH = 164
Na_2S = 67

Na_2CO_3 = 16
NAOH = 2
Na_2S = 5

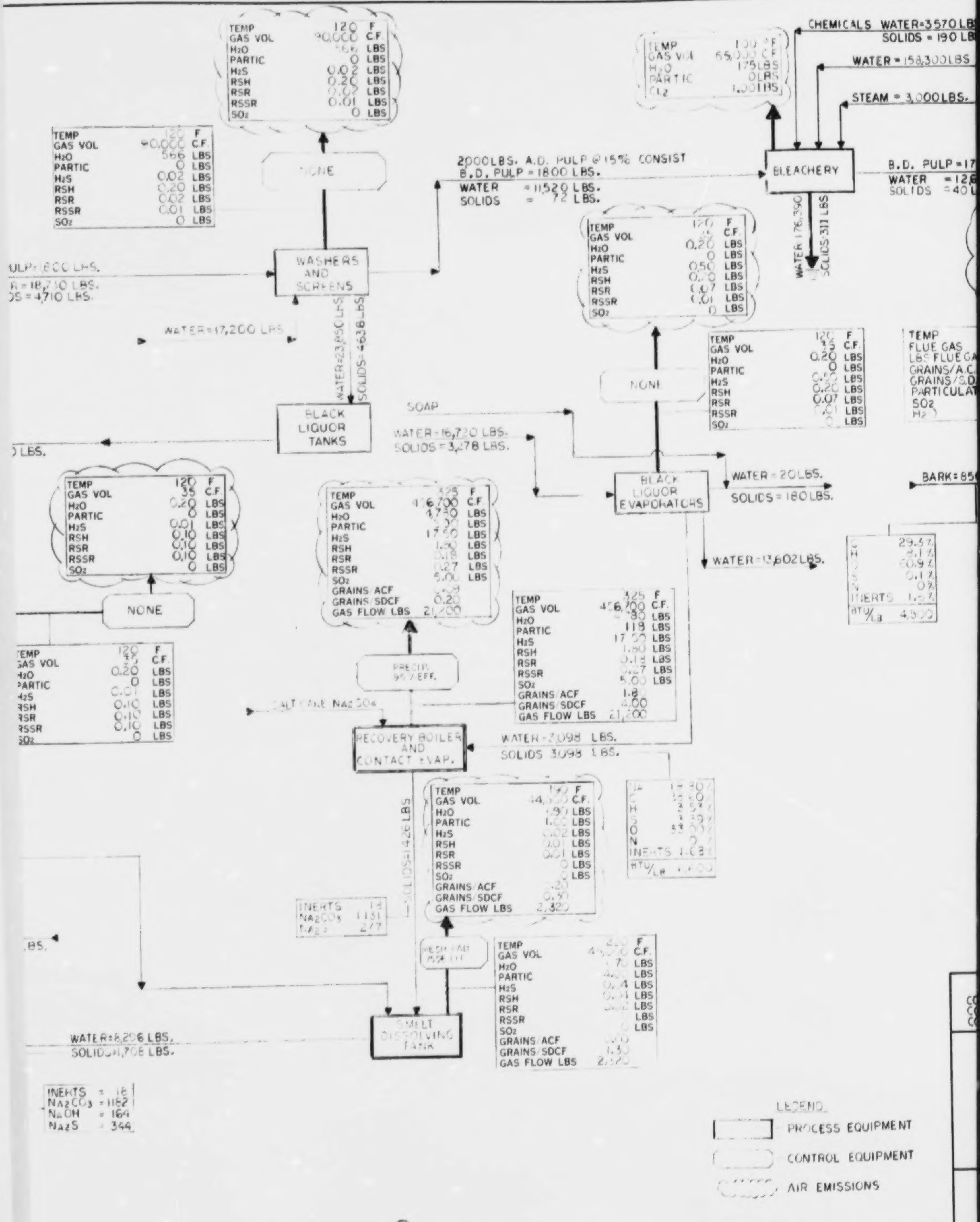
INERTS = 3
Na_2CO_3 = 1165
NAOH = 162
Na_2S = 339

CAO = 13.5
INERTS = 1.5

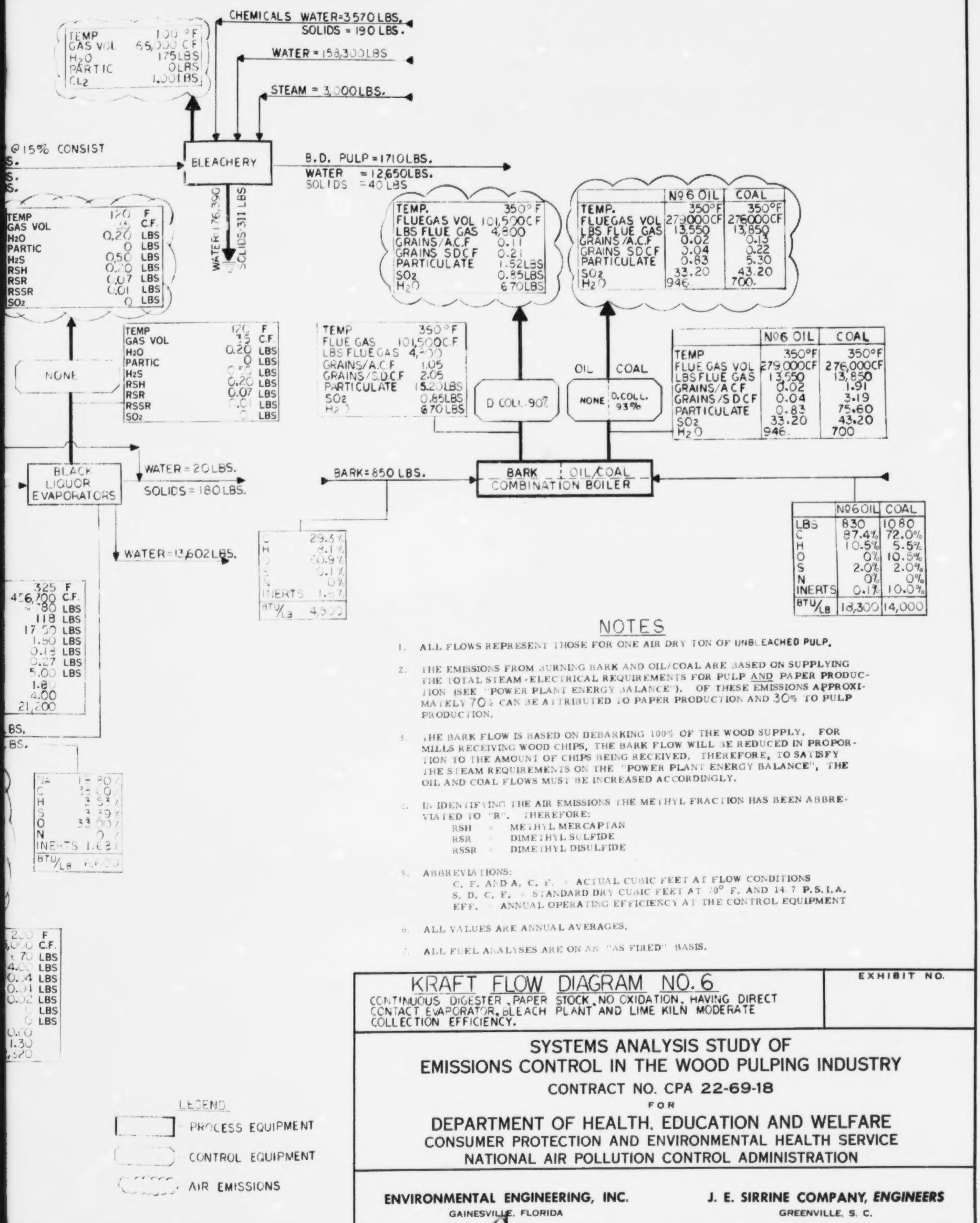
$CaCO_3$ = 14.5
INERTS = 3.5
Na_2CO_3 = 2.5
NAOH = 11.5
Na_2S = 4.5

INERTS = 15
Na_2CO_3 = 0.5
NAOH = TRACE
Na_2S = 0.5

A


TEMP 120 F
GAS VOL 90,000 C.F.
H2O 566 LBS
PARTIC 0 LBS
H2S 0.02 LBS
RSH 0.20 LBS
RSR 0.02 LBS
RSSR 0.01 LBS
SO2 0 LBS
NONE
WASHERS AND SCREENS
PULP=1800 LBS.
R=18,730 LBS.
DS=4,710 LBS.
WATER=17,200 LBS
WATER=23,850 LBS
SOLIDS=4636 LBS
BLACK LIQUOR TANKS
D LBS.
2000 LBS. A.D. PULP @ 15% CONSIST
B.D. PULP = 1800 LBS.
WATER = 11,520 LBS.
SOLIDS = 72 LBS.
CHEMICALS WATER=3570 LBS
SOLIDS = 190 LB
WATER = 158,300 LBS
STEAM = 3,000 LBS.
TEMP 100 F
GAS VOL 65,000 C.F.
H2O 175 LBS
PARTIC 0 LBS
CL2 1,001 LBS
BLEACHERY
B.D. PULP = 17
WATER = 12
SOLIDS = 40 L
WATER 176,390
SOLIDS 311 LBS
TEMP 120 F
GAS VOL 35 C.F.
H2O 0.20 LBS
PARTIC 0 LBS
H2S 0.50 LBS
RSH 0.20 LBS
RSR 0.07 LBS
RSSR 0.01 LBS
SO2 0 LBS
NONE
SOAP
WATER=16,720 LBS.
SOLIDS=3,278 LBS.
BLACK LIQUOR EVAPORATORS
WATER = 20 LBS.
SOLIDS = 180 LBS.
WATER = 13,602 LBS.
TEMP
FLUE GAS
LBS FLUE GA
GRAINS/A.C.
GRAINS/S.D.
PARTICULAT
SO2
H2O
BARK=85
C 29.3%
H 3.1%
O 60.9%
S 0.1%
N 0%
INERTS 1.5%
BTU/LB 4,500
TEMP 120 F
GAS VOL 35 C.F.
H2O 0.20 LBS
PARTIC 0 LBS
H2S 0.01 LBS
RSH 0.10 LBS
RSR 0.10 LBS
RSSR 0.10 LBS
SO2 0 LBS
NONE
TEMP 325 F
GAS VOL 426,700 C.F.
H2O 4,750 LBS
PARTIC 2.00 LBS
H2S 17.50 LBS
RSH 1.50 LBS
RSR 0.18 LBS
RSSR 0.27 LBS
SO2 5.00 LBS
GRAINS/ACF 0.09
GRAINS/SDCF 0.20
GAS FLOW LBS 21,200
PRECIP. 95% EFF.
SALT CAKE Na2SO4
RECOVERY BOILER AND CONTACT EVAP.
TEMP 325 F
GAS VOL 426,700 C.F.
H2O 4,780 LBS
PARTIC 118 LBS
H2S 17.50 LBS
RSH 1.50 LBS
RSR 0.18 LBS
RSSR 0.27 LBS
SO2 5.00 LBS
GRAINS/ACF 1.80
GRAINS/SDCF 4.00
GAS FLOW LBS 21,200
WATER=7,098 LBS.
SOLIDS 3,098 LBS.
Na 19.80%
C 34.60%
H 3.54%
S 3.35%
O 33.00%
N 0%
INERTS 1.63%
BTU/LB 6,600
SOLIDS=1426 LBS
TEMP 190 F
GAS VOL 44,000 C.F.
H2O 490 LBS
PARTIC 1.00 LBS
H2S 0.02 LBS
RSH 0.01 LBS
RSR 0.01 LBS
RSSR 0 LBS
SO2 0 LBS
GRAINS/ACF 0.20
GRAINS/SDCF 0.30
GAS FLOW LBS 2,320
INERTS 18
Na2CO3 1131
Na2S 277
SCRUBBER 75% EFF.
TEMP 200 F
GAS VOL 45,000 C.F.
H2O 470 LBS
PARTIC 4.00 LBS
H2S 0.04 LBS
RSH 0.04 LBS
RSR 0.02 LBS
RSSR 0 LBS
SO2 0 LBS
GRAINS/ACF 0.80
GRAINS/SDCF 1.30
GAS FLOW LBS 2,320
LBS.
SMELT DISSOLVING TANK
WATER=8,206 LBS.
SOLIDS=1,706 LBS.
INERTS = 18
Na2CO3 = 1182
NaOH = 164
Na2S = 344
TEMP 120 F
GAS VOL 35 C.F.
H2O 0.20 LBS
PARTIC 0 LBS
H2S 0.01 LBS
RSH 0.10 LBS
RSR 0.10 LBS
RSSR 0.10 LBS
SO2 0 LBS
LEGEND
PROCESS EQUIPMENT
CONTROL EQUIPMENT
AIR EMISSIONS

B


TEMP 100 °F
GAS VOL 55,000 CF
H_2O 175 LBS
PARTIC 0 LBS
CL_2 1,000 LBS
CHEMICALS WATER = 3570 LBS.
SOLIDS = 190 LBS.
WATER = 158,300 LBS
STEAM = 3,000 LBS.
@ 15% CONSIST
BLEACHERY
B.D. PULP = 1710 LBS.
WATER = 12,650 LBS.
SOLIDS = 40 LBS
WATER-176,390
SOLIDS-311 LBS
TEMP 120 F
GAS VOL 35 C.F.
H_2O 0.20 LBS
PARTIC 0 LBS
H_2S 0.50 LBS
RSH 0.20 LBS
RSR 0.07 LBS
RSSR 0.01 LBS
SO_2 0 LBS
NONE
TEMP 120 F
GAS VOL 35 C.F.
H_2O 0.20 LBS
PARTIC 0 LBS
H_2S 0.50 LBS
RSH 0.20 LBS
RSR 0.07 LBS
RSSR 0.01 LBS
SO_2 0 LBS
BLACK LIQUOR EVAPORATORS
WATER = 20 LBS.
SOLIDS = 180 LBS.
WATER = 13,602 LBS.
TEMP. 350 °F
FLUEGAS VOL 101,500 CF
LBS FLUE GAS 4,800
GRAINS/A.C.F 0.11
GRAINS SDCF 0.21
PARTICULATE 1.52 LBS
SO_2 0.85 LBS
H_2O 670 LBS
TEMP 350 °F
FLUE GAS 101,500 C.F
LBS FLUE GAS 4,800
GRAINS/A.C.F 1.05
GRAINS/S.D.C.F 2.05
PARTICULATE 15.20 LBS
SO_2 0.85 LBS
H_2O 670 LBS
D COLL. 90%
OIL
COAL
NONE
D.COLL. 93%
Nº6 OIL COAL
TEMP. 350°F 350°F
FLUEGAS VOL 279000 CF 276000 CF
LBS FLUE GAS 13,550 13,850
GRAINS/A.C.F 0.02 0.13
GRAINS SDCF 0.04 0.22
PARTICULATE 0.83 5.30
SO_2 33.20 43.20
H_2O 946. 700.
Nº6 OIL COAL
TEMP 350°F 350°F
FLUE GAS VOL 279,000 CF 276,000 CF
LBS FLUE GAS 13,550 13,850
GRAINS/A.C.F 0.02 1.91
GRAINS/S.D.C.F 0.04 3.19
PARTICULATE 0.83 75.60
SO_2 33.20 43.20
H_2O 946. 700
BARK = 850 LBS.
BARK OIL/COAL COMBINATION BOILER
C 29.3%
H 8.1%
O 60.9%
S 0.1%
N 0%
INERTS 1.6%
BTU/LB 4,500
Nº6 OIL COAL
LBS 830 1080
C 87.4% 72.0%
H 10.5% 5.5%
O 0% 10.5%
S 2.0% 2.0%
N 0% 0%
INERTS 0.1% 10.0%
BTU/LB 18,300 14,000
325 F
456,700 C.F.
780 LBS
118 LBS
17.50 LBS
1.50 LBS
0.18 LBS
0.27 LBS
5.00 LBS
1.80
4.00
21,200
LBS.
LBS.
NA 19.80%
C 35.00%
H 3.55%
S 3.59%
O 33.00%
N 0%
INERTS 1.68%
BTU/LB 6,600
200 F
000 C.F.
70 LBS
4.00 LBS
0.04 LBS
0.01 LBS
0.02 LBS
0 LBS
0 LBS
0.00
1.30
320
LEGEND
PROCESS EQUIPMENT
CONTROL EQUIPMENT
AIR EMISSIONS

NOTES

1. ALL FLOWS REPRESENT THOSE FOR ONE AIR DRY TON OF UNBLEACHED PULP.
2. THE EMISSIONS FROM BURNING BARK AND OIL/COAL ARE BASED ON SUPPLYING THE TOTAL STEAM-ELECTRICAL REQUIREMENTS FOR PULP AND PAPER PRODUCTION (SEE "POWER PLANT ENERGY BALANCE"). OF THESE EMISSIONS APPROXIMATELY 70% CAN BE ATTRIBUTED TO PAPER PRODUCTION AND 30% TO PULP PRODUCTION.
3. THE BARK FLOW IS BASED ON DEBARKING 100% OF THE WOOD SUPPLY. FOR MILLS RECEIVING WOOD CHIPS, THE BARK FLOW WILL BE REDUCED IN PROPORTION TO THE AMOUNT OF CHIPS BEING RECEIVED. THEREFORE, TO SATISFY THE STEAM REQUIREMENTS ON THE "POWER PLANT ENERGY BALANCE", THE OIL AND COAL FLOWS MUST BE INCREASED ACCORDINGLY.
4. IN IDENTIFYING THE AIR EMISSIONS THE METHYL FRACTION HAS BEEN ABBREVIATED TO "R". THEREFORE:
   - RSH = METHYL MERCAPTAN
   - RSR = DIMETHYL SULFIDE
   - RSSR = DIMETHYL DISULFIDE
5. ABBREVIATIONS:
   - C. F. AND A. C. F. = ACTUAL CUBIC FEET AT FLOW CONDITIONS
   - S. D. C. F. = STANDARD DRY CUBIC FEET AT 70° F. AND 14.7 P.S.I.A.
   - EFF. = ANNUAL OPERATING EFFICIENCY AT THE CONTROL EQUIPMENT
6. ALL VALUES ARE ANNUAL AVERAGES.
7. ALL FUEL ANALYSES ARE ON AN "AS FIRED" BASIS.

KRAFT FLOW DIAGRAM NO. 6

CONTINUOUS DIGESTER, PAPER STOCK, NO OXIDATION, HAVING DIRECT CONTACT EVAPORATOR, BLEACH PLANT AND LIME KILN MODERATE COLLECTION EFFICIENCY.

EXHIBIT NO.

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CONSUMER PROTECTION AND ENVIRONMENTAL HEALTH SERVICE
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

ENVIRONMENTAL ENGINEERING, INC.
GAINESVILLE, FLORIDA

J. E. SIRRINE COMPANY, ENGINEERS
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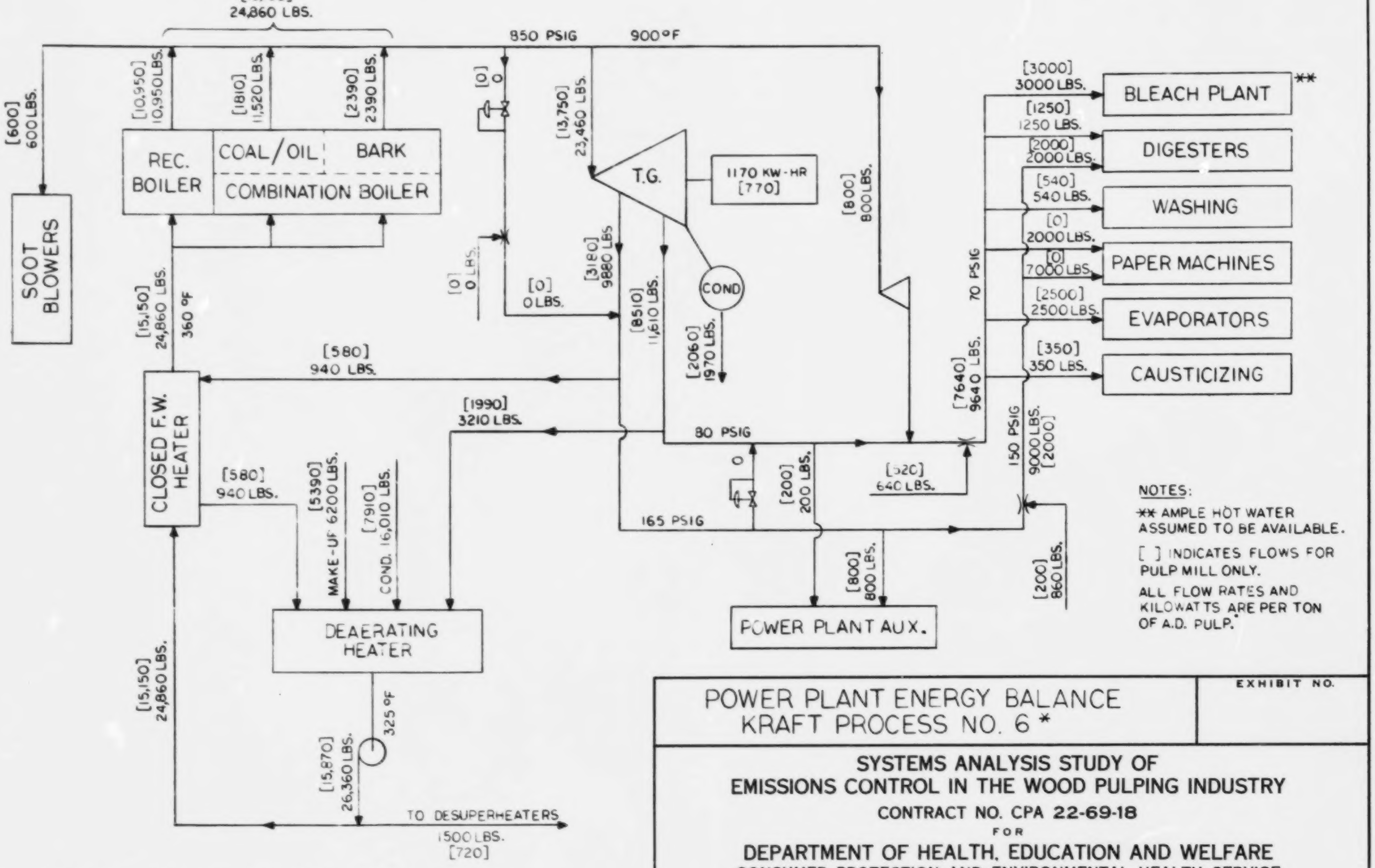
[15,150]
24,860 LBS.
850 PSIG
900°F
[600]
600 LBS.
SOOT BLOWERS
[10,950]
10,950 LBS.
[1810]
11,520 LBS.
[2390]
2390 LBS.
REC. BOILER
COAL/OIL
BARK
COMBINATION BOILER
[15,150]
24,860 LBS.
360 °F
[0]
0
[0]
0 LBS.
[0]
0 LBS.
[13,750]
23,460 LBS.
T.G.
1170 KW-HR
[770]
[3180]
9880 LBS.
[8510]
11,610 LBS.
COND
[2060]
1970 LBS.
[800]
800 LBS.
[580]
940 LBS.
[1990]
3210 LBS.
80 PSIG
0
[200]
200 LBS.
[520]
640 LBS.
165 PSIG
[800]
800 LBS.
POWER PLANT AUX.
CLOSED F.W. HEATER
[580]
940 LBS.
[5390]
MAKE-UP 6200 LBS.
[7910]
COND. 16,010 LBS.
DEAERATING HEATER
325 °F
[15,870]
26,360 LBS.
[15,150]
24,860 LBS.
TO DESUPERHEATERS
1500 LBS.
[720]
[7640]
9640 LBS.
70 PSIG
150 PSIG
9000 LBS.
[2000]
[200]
860 LBS.
[3000]
3000 LBS.
BLEACH PLANT **
[1250]
1250 LBS.
DIGESTERS
[2000]
2000 LBS.
[540]
540 LBS.
WASHING
[0]
2000 LBS.
[0]
7000 LBS.
PAPER MACHINES
[2500]
2500 LBS.
EVAPORATORS
[350]
350 LBS.
CAUSTICIZING

NOTES:

** AMPLE HOT WATER ASSUMED TO BE AVAILABLE.

[ ] INDICATES FLOWS FOR PULP MILL ONLY.

ALL FLOW RATES AND KILOWATTS ARE PER TON OF A.D. PULP.

* CONTINUOUS DIGESTERS, PAPER STOCK, NO OXIDATION, DIRECT CONTACT EVAPORATORS, BLEACH PLANT, MODERATE LIME KILN COLLECTION EFFICIENCY

POWER PLANT ENERGY BALANCE KRAFT PROCESS NO. 6 *

EXHIBIT NO.

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NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

ENVIRONMENTAL ENGINEERING, INC.
GAINESVILLE, FLORIDA

J. E. SIRRINE COMPANY, *ENGINEERS*
GREENVILLE, S. C.

KRAFT DIAGRAM NO. 7

TYPICAL LOCATION

Northeastern United States.

TYPICAL AGE OF EQUIPMENT

Ten to 15 years.

GENERAL

This flow diagram is based on production of bleached stock with cooking done in batch digesters. This flow diagram depicts the use of a Venturi scrubber rather than the commonly used electrostatic precipitator. A mesh pad has been assumed in the vent from the smelt dissolving tank. There are perhaps 15 to 25 mills in the U. S. that are illustrated by this flow diagram.

EMISSIONS

The following emissions have been selected as representative, based on the range of emissions presented in Chapter 4.

POUNDS PER AIR DRY TON OF UNBLEACHED PULP

LOCATION	H_2S	RSH, RSR, RSSR	SO_2	Particulate
Accumulator	0.12	2.97	0	0
Washers and Screens	0.01	0.22	0	0
M. E. Evaporators	0.50	0.41	0	0
Recovery Boiler After Venturi Scrubber	8.0	1.4	2.3	47.6
Smelt Dissolving Tank	0.02	0.02	0	1.00
Slaker	0	0	0	Unknown
Lime Kiln	1.0	0.6	Trace	55
Lime Kiln After Scrubber	0.4	0.23	Trace	11.0
Turpentine Condenser	0.01	0.50	0	0

KRAFT DIAGRAM NO. 7

(Continued)

ASSUMPTIONS

The following assumptions have been made in developing the flow diagram:

A. PULP MILL

1. Pulp Yield = 45 Percent.
2. Cooking liquor charge per air dry ton of brown stock equals 6,996 pounds of which 996 pounds are chemical soids. Sulfidity equals 25 percent.
3. Black liquor solids from first stage washer equals 15.5 percent and from M. E. Evaporators equals 50 percent solids.

B. BLEACH PLANT

1. Assumed a five stage bleach plant with a chlorine emission of one pound per air dry ton of pulp.

C. RECOVERY

1. Unit operating at 15 percent excess air at economizer outlet.
2. Particulate matter of 8 grains per SDCF leaving the economizer.
3. Steam shatter jets utilized on smelt spouts.
4. Design efficiency of 90 percent for Venturi scrubber with an annual average operating efficiency of 80 percent.
5. Efficiency of 80 percent on particulate removal in the scrubber.

D. CAUSTICIZING

1. Lime Kiln scrubber efficiency assumed at 80 percent on lime and 60 percent efficiency on soda.

KRAFT DIAGRAM NO. 7
(Continued)

E. POWER PLANT

1. Excess air: 30 percent for bark, 10 percent for oil and 20 percent for coal.

2. Bark burning equipment is a low set spreader stoker.

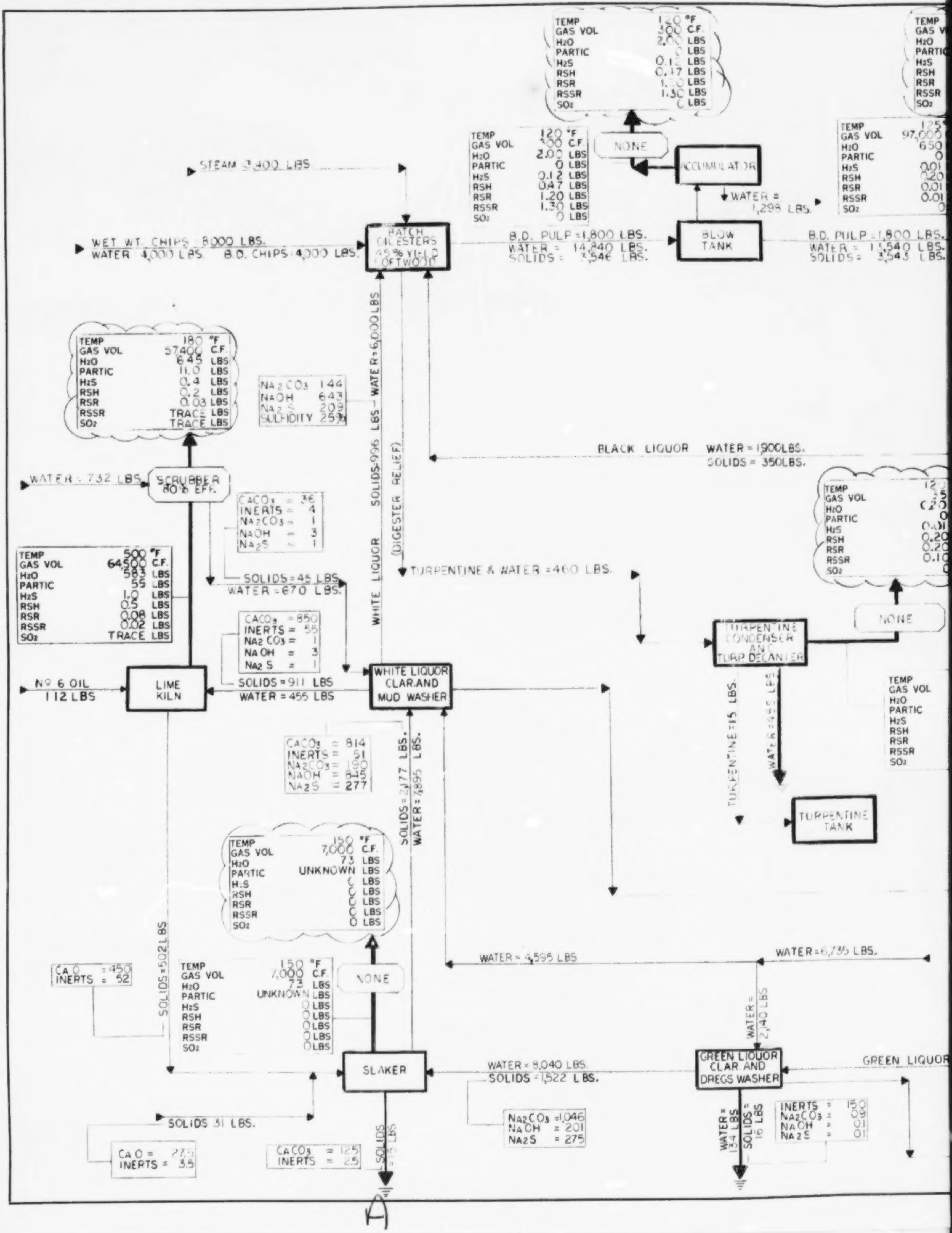
3. Reinjection from dust collector: 50 percent for bark and 0 percent for coal.

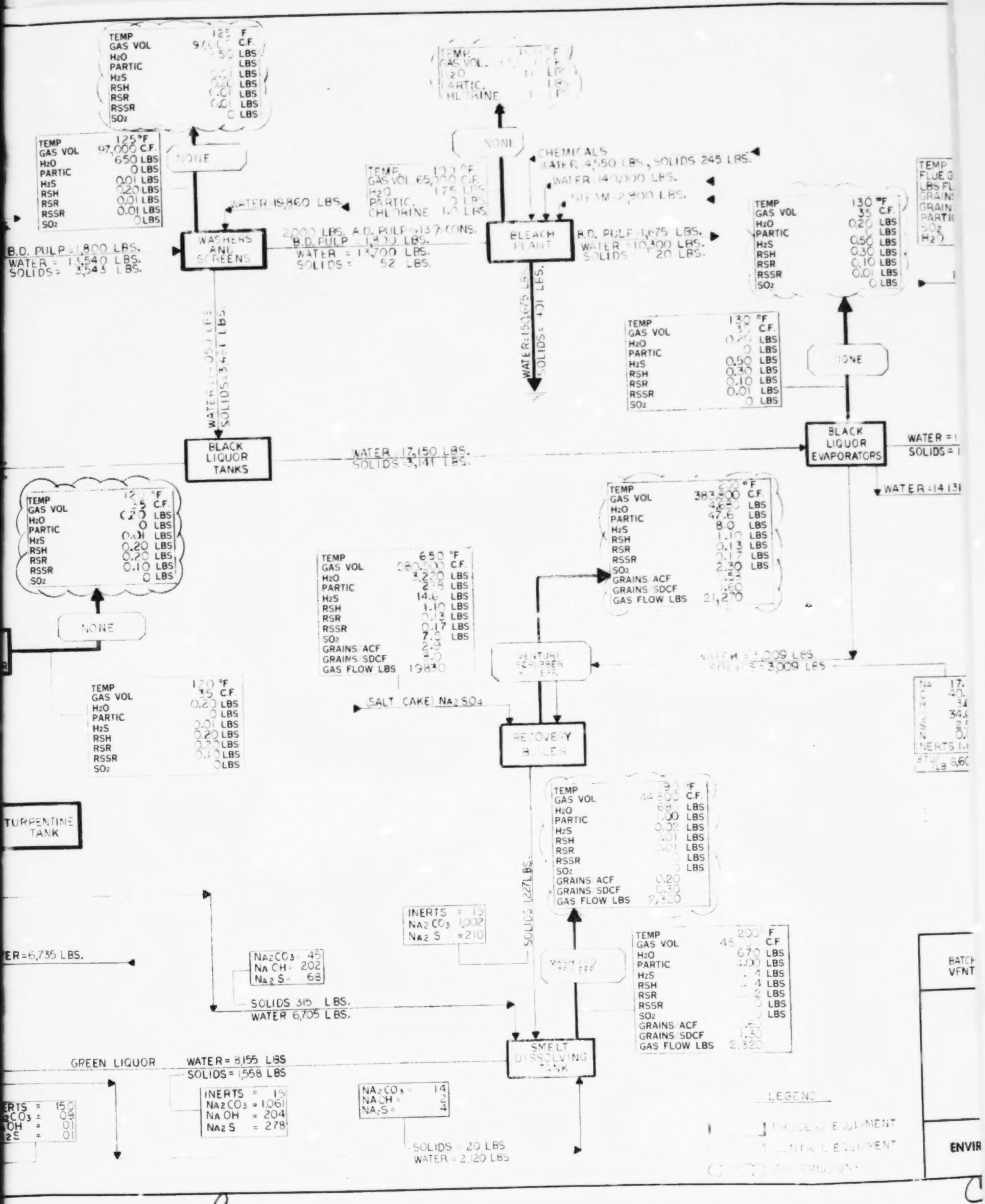
4. Design efficiencies for dust collector: 80 percent on bark and 90 percent on coal with annual operating efficiencies of 78 percent and 88 percent, respectively.

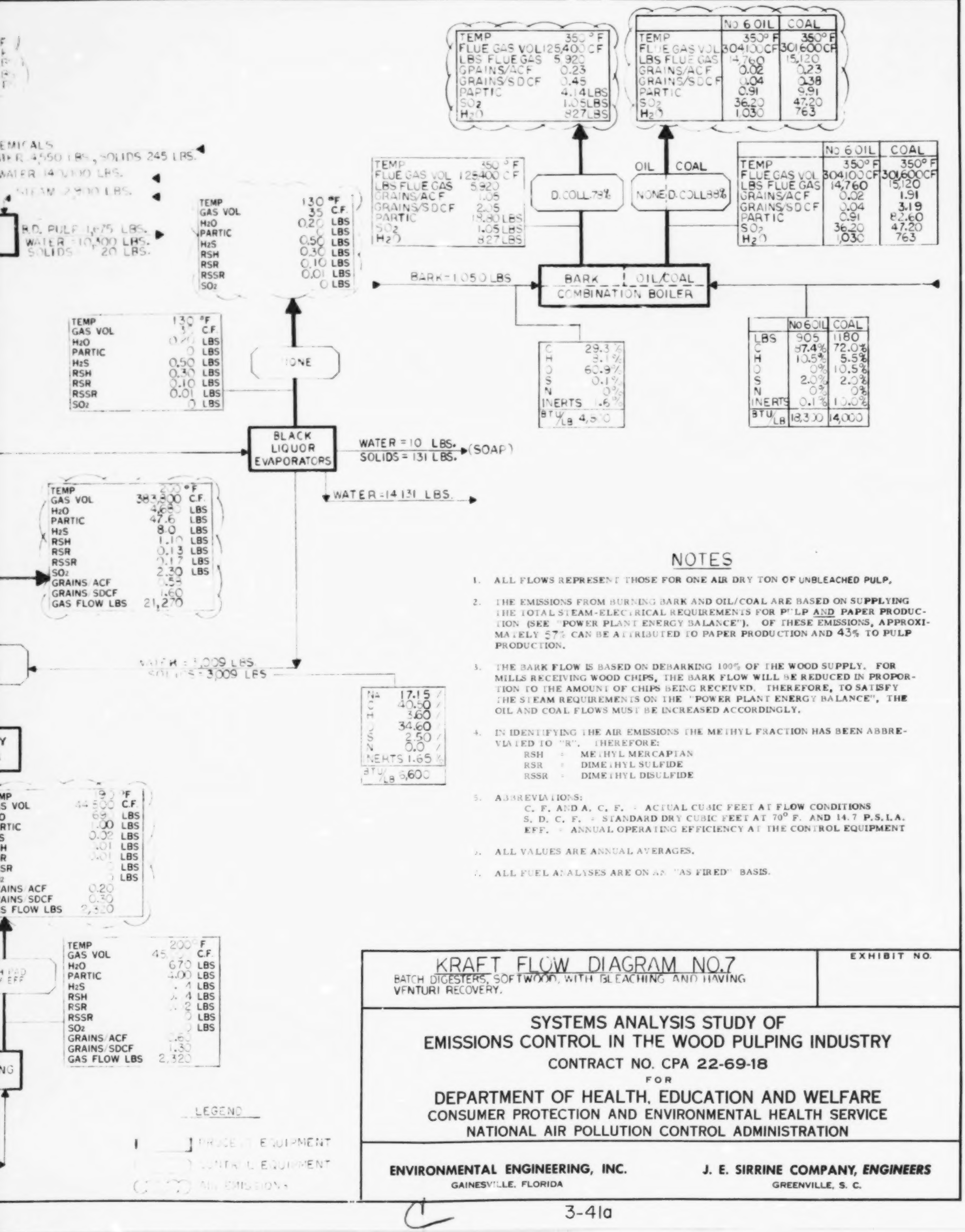
5. Bark is burned 24 hours per day at a controlled rate.

6. Unburned combustible in refuse leaving the dust collector is 40 percent.

7. Coal firing based on pulverized coal.


TEMP 120 °F
GAS VOL 300 C.F.
H2O 2.00 LBS
PARTIC 0 LBS
H2S 0.12 LBS
RSH 0.47 LBS
RSR 1.20 LBS
RSSR 1.30 LBS
SO2 0 LBS
NONE
ACCUMULATOR
WATER = 1,298 LBS.
STEAM 3,400 LBS
WET WT. CHIPS 8000 LBS.
WATER 4,000 LBS. B.D. CHIPS 4,000 LBS.
BATCH DIGESTERS 45% YIELD SOFTWOOD
B.D. PULP = 1,800 LBS.
WATER = 14,840 LBS.
SOLIDS = 3,546 LBS.
BLOW TANK
B.D. PULP = 1,800 LBS.
WATER = 13,540 LBS.
SOLIDS = 3,543 LBS.
TEMP 125
GAS VOL 97,000
H2O 650
PARTIC 0
H2S 0.01
RSH 0.20
RSR 0.01
RSSR 0.01
SO2 0
TEMP 180 °F
GAS VOL 57400 C.F.
H2O 645 LBS
PARTIC 11.0 LBS
H2S 0.4 LBS
RSH 0.2 LBS
RSR 0.03 LBS
RSSR TRACE LBS
SO2 TRACE LBS
NA2CO3 144
NAOH 643
NA2S 209
SULFIDITY 25%
WHITE LIQUOR SOLIDS-996 LBS - WATER = 6,000 LBS
(DIGESTER RELIEF)
BLACK LIQUOR WATER = 1900 LBS.
SOLIDS = 350 LBS.
WATER = 732 LBS.
SCRUBBER 80% EFF.
CACO3 = 36
INERTS = 4
NA2CO3 = 1
NAOH = 3
NA2S = 1
SOLIDS = 45 LBS.
WATER = 670 LBS.
TURPENTINE & WATER = 460 LBS.
TEMP 500 °F
GAS VOL 64,500 C.F.
H2O 583 LBS
PARTIC 55 LBS
H2S 1.0 LBS
RSH 0.5 LBS
RSR 0.08 LBS
RSSR 0.02 LBS
SO2 TRACE LBS
CACO3 = 850
INERTS = 55
NA2CO3 = 1
NAOH = 3
NA2S = 1
SOLIDS = 911 LBS
WATER = 455 LBS
Nº 6 OIL 112 LBS
LIME KILN
WHITE LIQUOR CLAR. AND MUD WASHER
TURPENTINE CONDENSER AND TURP DECANTER
NONE
TEMP 120
GAS VOL
H2O
PARTIC 0.01
H2S 0.20
RSH 0.20
RSR 0.10
RSSR 0
SO2
TEMP
GAS VOL
H2O
PARTIC
H2S
RSH
RSR
RSSR
SO2
TURPENTINE = 15 LBS.
WATER = 445 LBS.
TURPENTINE TANK
CACO3 = 814
INERTS = 51
NA2CO3 = 190
NAOH = 845
NA2S = 277
SOLIDS = 2177 LBS.
WATER = 7,895 LBS.
TEMP 150 °F
GAS VOL 7,000 C.F.
H2O 73 LBS
PARTIC UNKNOWN LBS
H2S 0 LBS
RSH 0 LBS
RSR 0 LBS
RSSR 0 LBS
SO2 0 LBS
SOLIDS = 502 LBS
CA O = 450
INERTS = 52
TEMP 150 °F
GAS VOL 7,000 C.F.
H2O 73 LBS
PARTIC UNKNOWN LBS
H2S 0 LBS
RSH 0 LBS
RSR 0 LBS
RSSR 0 LBS
SO2 0 LBS
NONE
WATER = 4,595 LBS
WATER = 6,735 LBS.
WATER = 2,140 LBS
SLAKER
WATER = 8,040 LBS.
SOLIDS = 1,522 LBS.
GREEN LIQUOR CLAR. AND DREGS WASHER
GREEN LIQUOR
SOLIDS 31 LBS.
CA O = 27.5
INERTS = 3.5
CACO3 = 12.5
INERTS = 2.5
SOLIDS = 15 LBS
NA2CO3 = 1,046
NAOH = 201
NA2S = 275
WATER = 134 LBS
SOLIDS = 16 LBS
INERTS = 15.0
NA2CO3 = .09
NAOH = .01
NA2S = .01


TEMP 125 F
GAS VOL 97,000 C.F.
H2O 650 LBS
PARTIC LBS
H2S 0.01 LBS
RSH 0.20 LBS
RSR 0.01 LBS
RSSR 0.01 LBS
SO2 0 LBS
TEMP 125°F
GAS VOL 97,000 C.F.
H2O 650 LBS
PARTIC 0 LBS
H2S 0.01 LBS
RSH 0.20 LBS
RSR 0.01 LBS
RSSR 0.01 LBS
SO2 0 LBS
NONE
WATER 19,860 LBS.
B.D. PULP = 1,800 LBS.
WATER = 13,540 LBS.
SOLIDS = 3,543 LBS.
WASHERS AND SCREENS
2,000 LBS. A.D. PULP @ 13% CONS.
B.D. PULP = 1,800 LBS.
WATER = 13,700 LBS.
SOLIDS = 52 LBS.
TEMP. °F
GAS VOL. C.F.
H2O LBS
PARTIC. LBS
CHLORINE LBS
NONE
TEMP. 100 °F
GAS VOL. 65,000 C.F.
H2O 175 LBS
PARTIC. 0 LBS
CHLORINE 1.0 LBS
CHEMICALS
WATER 4,550 LBS., SOLIDS 245 LBS.
WATER 140,000 LBS.
STEAM 2,900 LBS.
BLEACH PLANT
B.D. PULP 1,675 LBS.
WATER = 10,300 LBS.
SOLIDS = 20 LBS.
WATER = 150,675 LBS.
SOLIDS = 401 LBS.
WATER = LBS
SOLIDS = 3,4 1 LBS.
TEMP 130 °F
GAS VOL 35 C.F.
H2O 0.20 LBS
PARTIC 0 LBS
H2S 0.50 LBS
RSH 0.30 LBS
RSR 0.10 LBS
RSSR 0.01 LBS
SO2 0 LBS
TEMP
FLUE G
LBS FL
GRAINS
GRAIN
PARTIC
SO2
H2O
TEMP 130 °F
GAS VOL 35 C.F.
H2O 0.20 LBS
PARTIC 0 LBS
H2S 0.50 LBS
RSH 0.30 LBS
RSR 0.10 LBS
RSSR 0.01 LBS
SO2 0 LBS
NONE
BLACK LIQUOR TANKS
WATER = 17,150 LBS.
SOLIDS = 3,141 LBS.
BLACK LIQUOR EVAPORATORS
WATER = 1
SOLIDS = 1
WATER = 14,13
TEMP 120 °F
GAS VOL 35 C.F.
H2O 0.20 LBS
PARTIC 0 LBS
H2S 0.01 LBS
RSH 0.20 LBS
RSR 0.20 LBS
RSSR 0.10 LBS
SO2 0 LBS
NONE
TEMP 200 °F
GAS VOL 383,800 C.F.
H2O 4,680 LBS
PARTIC 47.6 LBS
H2S 8.0 LBS
RSH 1.10 LBS
RSR 0.13 LBS
RSSR 0.17 LBS
SO2 2.30 LBS
GRAINS/ACF .53
GRAINS/SDCF 1.60
GAS FLOW LBS 21,270
TEMP 650 °F
GAS VOL 580,500 C.F.
H2O 3,220 LBS
PARTIC 2 8 LBS
H2S 14.6 LBS
RSH 1.10 LBS
RSR 0.13 LBS
RSSR 0.17 LBS
SO2 7.5 LBS
GRAINS/ACF 2.9
GRAINS/SDCF 8.0
GAS FLOW LBS 19830
VENTURI SCRUBBER % EFF.
WATER = 3,009 LBS
SOLIDS = 3,009 LBS
TEMP 120 °F
GAS VOL 35 C.F.
H2O 0.20 LBS
PARTIC 0 LBS
H2S 0.01 LBS
RSH 0.20 LBS
RSR 0.20 LBS
RSSR 0.10 LBS
SO2 0 LBS
(SALT CAKE) NA2 SO4
RECOVERY BOILER
NA 17.
C 40.
H 3.
O 34.
S 2.
N 0.
INERTS 1.
BTU/LB 6,60
TURPENTINE TANK
TEMP 195 °F
GAS VOL 44,500 C.F.
H2O 69 LBS
PARTIC 1.00 LBS
H2S 0.02 LBS
RSH 0.01 LBS
RSR 0.01 LBS
RSSR 0 LBS
SO2 0 LBS
GRAINS/ACF 0.20
GRAINS/SDCF 0.30
GAS FLOW LBS 2,320
SOLIDS 1,227 LBS.
INERTS = 15
NA2 CO3 1,002
NA2 S = 210
ER = 6,735 LBS.
NA2CO3 = 45
NA OH = 202
NA2 S = 68
SOLIDS 315 LBS.
WATER 6,705 LBS.
MESH PAD % EFF
TEMP 200 °F
GAS VOL 45, C.F.
H2O 670 LBS
PARTIC 4.00 LBS
H2S 4 LBS
RSH 4 LBS
RSR 2 LBS
RSSR 0 LBS
SO2 0 LBS
GRAINS/ACF .50
GRAINS/SDCF 1.30
GAS FLOW LBS 2,320
BATCH VENT
SMELT DISSOLVING TANK
GREEN LIQUOR
WATER = 8,155 LBS
SOLIDS = 1,558 LBS
ERTS = 150
2CO3 = 09
OH = 01
2S = 01
INERTS = 15
NA2 CO3 = 1,061
NA OH = 204
NA2 S = 278
NA2CO3 = 14
NA OH = 2
NA2S = 4
SOLIDS = 20 LBS
WATER = 2,120 LBS
LEGEND
PROCESS EQUIPMENT
CONTROL EQUIPMENT
AIR EMISSIONS
ENVIR
B
C


EMICALS
TER 4,550 LBS., SOLIDS 245 LBS.
WATER 140,000 LBS.
STEAM 2,900 LBS.
B.D. PULP 1,675 LBS.
WATER 10,400 LBS.
SOLIDS 20 LBS.
TEMP 130 °F
GAS VOL 35 C.F.
H2O 0.20 LBS
PARTIC 0 LBS
H2S 0.50 LBS
RSH 0.30 LBS
RSR 0.10 LBS
RSSR 0.01 LBS
SO2 0 LBS
NONE
BLACK LIQUOR EVAPORATORS
WATER = 10 LBS.
SOLIDS = 131 LBS.
(SOAP)
WATER = 14,131 LBS.
TEMP 200 °F
GAS VOL 383,800 C.F.
H2O 4,680 LBS
PARTIC 47.6 LBS
H2S 8.0 LBS
RSH 1.10 LBS
RSR 0.13 LBS
RSSR 0.17 LBS
SO2 2.30 LBS
GRAINS/ACF 0.55
GRAINS/SDCF 1.60
GAS FLOW LBS 21,270
WATER = 3,009 LBS.
SOLIDS = 3,009 LBS.
NA 17.15 %
C 40.50 %
H 3.60 %
O 34.60 %
S 2.50 %
N 0.0 %
INERTS 1.65 %
BTU/LB 6,600
TEMP 190 °F
GAS VOL 44,500 C.F.
H2O 690 LBS
PARTIC 1.00 LBS
H2S 0.02 LBS
RSH 0.01 LBS
RSR 0.01 LBS
RSSR 0 LBS
SO2 0 LBS
GRAINS/ACF 0.20
GRAINS/SDCF 0.30
GAS FLOW LBS 2,320
TEMP 200 °F
GAS VOL 45,000 C.F.
H2O 670 LBS
PARTIC 4.00 LBS
H2S 0.4 LBS
RSH 0.4 LBS
RSR 0.2 LBS
RSSR 0 LBS
SO2 0 LBS
GRAINS/ACF 0.60
GRAINS/SDCF 1.30
GAS FLOW LBS 2,320
PAD EFF
NG
TEMP 350 °F
FLUE GAS VOL 125,400 CF
LBS FLUE GAS 5,920
GRAINS/ACF 0.23
GRAINS/SDCF 0.45
PARTIC 4.14 LBS
SO2 1.05 LBS
H2O 827 LBS
NO 6 OIL | COAL
TEMP 350° F | 350° F
FLUE GAS VOL 304,100 CF | 301,600 CF
LBS FLUE GAS 14,760 | 15,120
GRAINS/ACF 0.02 | 0.23
GRAINS/SDCF 0.04 | 0.38
PARTIC 0.91 | 9.91
SO2 36.20 | 47.20
H2O 1,030 | 763
TEMP 350 °F
FLUE GAS VOL 125,400 CF
LBS FLUE GAS 5,920
GRAINS/ACF 1.05
GRAINS/SDCF 2.05
PARTIC 18.80 LBS
SO2 1.05 LBS
H2O 827 LBS
D. COLL. 78%
OIL
COAL
NONE
D. COLL. 88%
NO 6 OIL | COAL
TEMP 350° F | 350° F
FLUE GAS VOL 304,100 CF | 301,600 CF
LBS FLUE GAS 14,760 | 15,120
GRAINS/ACF 0.02 | 1.91
GRAINS/SDCF 0.04 | 3.19
PARTIC 0.91 | 82.60
SO2 36.20 | 47.20
H2O 1,030 | 763
BARK = 1,050 LBS
BARK | OIL/COAL
COMBINATION BOILER
C 29.3 %
H 3.1 %
O 60.9 %
S 0.1 %
N 0 %
INERTS 1.6 %
BTU/LB 4,500
NO 6 OIL | COAL
LBS 905 | 1180
C 87.4% | 72.0%
H 10.5% | 5.5%
O 0% | 10.5%
S 2.0% | 2.0%
N 0% | 0%
INERTS 0.1% | 10.0%
BTU/LB 18,300 | 14,000
LEGEND
PROCESS EQUIPMENT
CONTROL EQUIPMENT
AIR EMISSIONS

NOTES

1. ALL FLOWS REPRESENT THOSE FOR ONE AIR DRY TON OF UNBLEACHED PULP.
2. THE EMISSIONS FROM BURNING BARK AND OIL/COAL ARE BASED ON SUPPLYING THE TOTAL STEAM-ELECTRICAL REQUIREMENTS FOR PULP AND PAPER PRODUCTION (SEE "POWER PLANT ENERGY BALANCE"). OF THESE EMISSIONS, APPROXIMATELY 57% CAN BE ATTRIBUTED TO PAPER PRODUCTION AND 43% TO PULP PRODUCTION.
3. THE BARK FLOW IS BASED ON DEBARKING 100% OF THE WOOD SUPPLY. FOR MILLS RECEIVING WOOD CHIPS, THE BARK FLOW WILL BE REDUCED IN PROPORTION TO THE AMOUNT OF CHIPS BEING RECEIVED. THEREFORE, TO SATISFY THE STEAM REQUIREMENTS ON THE "POWER PLANT ENERGY BALANCE", THE OIL AND COAL FLOWS MUST BE INCREASED ACCORDINGLY.
4. IN IDENTIFYING THE AIR EMISSIONS THE METHYL FRACTION HAS BEEN ABBREVIATED TO "R". THEREFORE:
   RSH = METHYL MERCAPTAN
   RSR = DIMETHYL SULFIDE
   RSSR = DIMETHYL DISULFIDE
5. ABBREVIATIONS:
   C. F. AND A. C. F. = ACTUAL CUBIC FEET AT FLOW CONDITIONS
   S. D. C. F. = STANDARD DRY CUBIC FEET AT 70° F. AND 14.7 P.S.I.A.
   EFF. = ANNUAL OPERATING EFFICIENCY AT THE CONTROL EQUIPMENT
6. ALL VALUES ARE ANNUAL AVERAGES.
7. ALL FUEL ANALYSES ARE ON AN "AS FIRED" BASIS.

KRAFT FLOW DIAGRAM NO.7

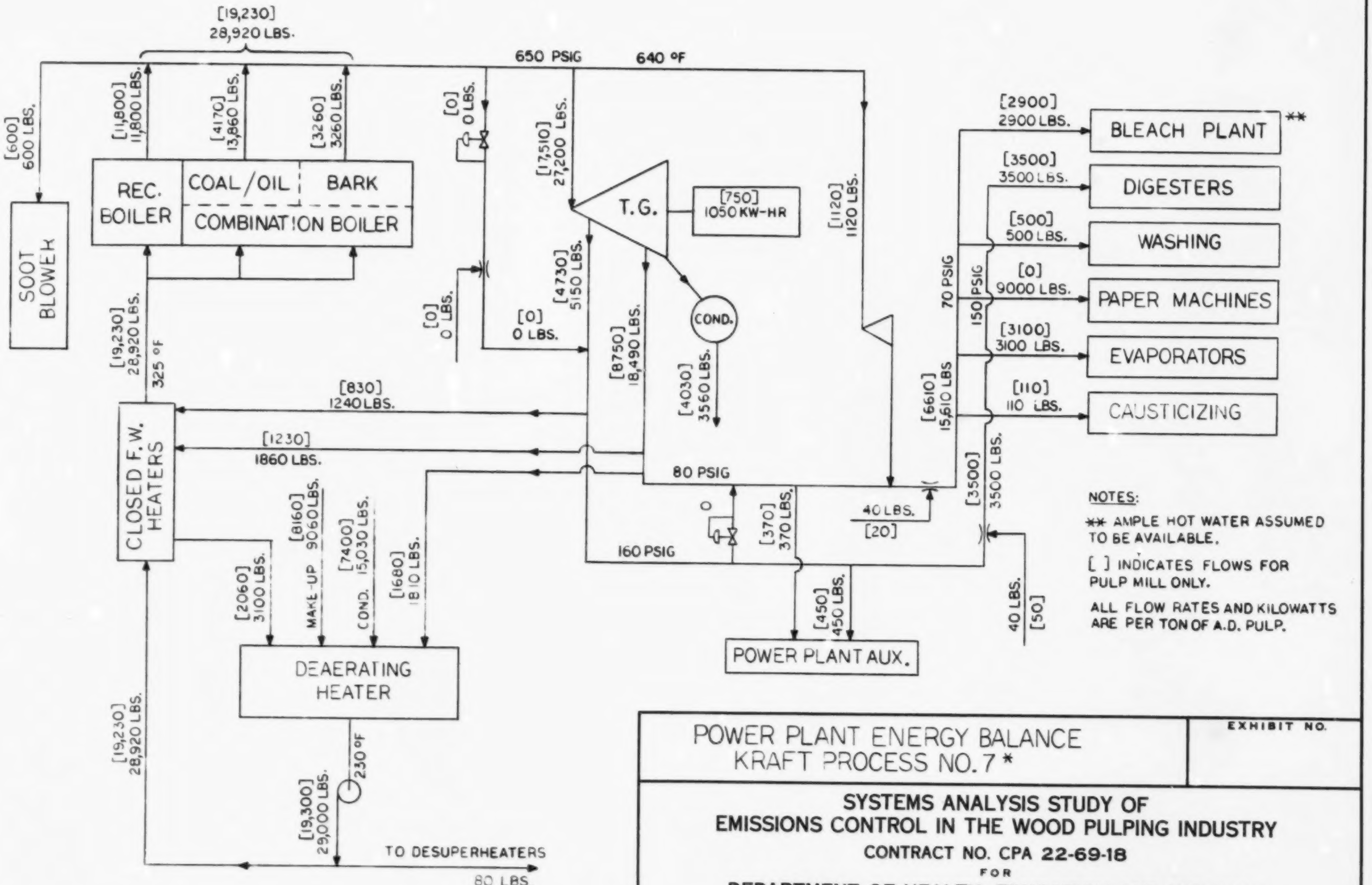
BATCH DIGESTERS, SOFTWOOD, WITH BLEACHING AND HAVING VENTURI RECOVERY.

EXHIBIT NO.

SYSTEMS ANALYSIS STUDY OF
EMISSIONS CONTROL IN THE WOOD PULPING INDUSTRY
CONTRACT NO. CPA 22-69-18
FOR
DEPARTMENT OF HEALTH, EDUCATION AND WELFARE
CONSUMER PROTECTION AND ENVIRONMENTAL HEALTH SERVICE
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

ENVIRONMENTAL ENGINEERING, INC.
GAINESVILLE, FLORIDA

J. E. SIRRINE COMPANY, *ENGINEERS*
GREENVILLE, S. C.


[19,230]
28,920 LBS.
650 PSIG
640 °F
[600]
600 LBS.
SOOT BLOWER
[11,800]
11,800 LBS.
[4170]
13,860 LBS.
[3260]
3260 LBS.
REC. BOILER
COAL/OIL
BARK
COMBINATION BOILER
[0]
0 LBS.
[0]
0 LBS.
[0]
0 LBS.
[17,510]
27,200 LBS.
T.G.
[750]
1050 KW-HR
[4730]
5150 LBS.
[8750]
18,490 LBS.
COND.
[4030]
3560 LBS.
[1120]
1120 LBS.
[19,230]
28,920 LBS.
325 °F
[830]
1240 LBS.
[1230]
1860 LBS.
CLOSED F.W. HEATERS
80 PSIG
160 PSIG
0
[370]
370 LBS.
40 LBS.
[20]
[6610]
15,610 LBS
70 PSIG
150 PSIG
[3500]
3500 LBS.
40 LBS.
[50]
[450]
450 LBS.
POWER PLANT AUX.
[2900]
2900 LBS.
BLEACH PLANT **
[3500]
3500 LBS.
DIGESTERS
[500]
500 LBS.
WASHING
[0]
9000 LBS.
PAPER MACHINES
[3100]
3100 LBS.
EVAPORATORS
[110]
110 LBS.
CAUSTICIZING
[2060]
3100 LBS.
[8160]
MAKE-UP 9060 LBS.
[7400]
COND. 15,030 LBS.
[1680]
1810 LBS.
DEAERATING HEATER
230 °F
[19,300]
29,000 LBS.
[19,230]
28,920 LBS.
TO DESUPERHEATERS
80 LBS.
[70]

NOTES:

** AMPLE HOT WATER ASSUMED TO BE AVAILABLE.

[ ] INDICATES FLOWS FOR PULP MILL ONLY.

ALL FLOW RATES AND KILOWATTS ARE PER TON OF A.D. PULP.

* BATCH DIGESTERS, NO BLACK LIQUOR OXIDATION, VENTURI RECOVERY, SOFTWOOD BLEACHING, LIME KILN MODERATE COLLECTION EFFICIENCY.

POWER PLANT ENERGY BALANCE
KRAFT PROCESS NO. 7 *

EXHIBIT NO.

SYSTEMS ANALYSIS STUDY OF
EMISSIONS CONTROL IN THE WOOD PULPING INDUSTRY
CONTRACT NO. CPA 22-69-18
FOR
DEPARTMENT OF HEALTH, EDUCATION AND WELFARE
CONSUMER PROTECTION AND ENVIRONMENTAL HEALTH SERVICE
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

ENVIRONMENTAL ENGINEERING, INC.
GAINESVILLE, FLORIDA

J. E. SIRRINE COMPANY, *ENGINEERS*
GREENVILLE, S. C.

KRAFT DIAGRAM NO. 8

TYPICAL LOCATION

Midwestern United States.

TYPICAL AGE OF EQUIPMENT

Under five years.

GENERAL

This flow diagram is based on production of bleached hardwood pulp in a continuous digester. It depicts a relatively new mill with a system for the treatment of noncondensible gases from the digester, multiple effect evaporators and the combined condensate from the multiple effect evaporators. There are perhaps 1 to 4 mills in the U. S. which are illustrated by this flow diagram.

EMISSIONS

The following emissions have been selected as representative, based on the range of emissionspresented in Chapter 4.

POUNDS PER AIR DRY TON OF UNBLEACHED PULP

LOCATION	H_2S	RSH RSR, RSSR	SO_2	Particulate
Washers and Screens	0.02	0.44	0	0
M. E. Evaporators	0.50	0.82	0	0
R. Boilers and D. C. Evaporators	17.6	0.36	7.10	120
After Precipitator	17.6	0.36	7.1	6.00
Smelt Dissolving Tank	0.04	0.06	0	4.0
Slaker	0	0	0	Unknown
Lime Kiln	1.0	0.58	Trace	57
After Lime Kiln Scrubber	0.2	0.11	Trace	0.50
Flashsteam Condenser	0.01	1.50	0	0

ASSUMPTIONS

The following assumptions have been made in developing the flow diagram:

A. PULP MILL

1. Pulp Yield = 46 Percent.

2. Cooking liquor charge per air dry ton of brown stock equals 6,935 pounds of which 972 pounds are chemical solids. Sulfidity equals 24 percent.

B. BLEACH PLANT

1. Assumed a five stage bleach plant with a chlorine emission of one pound per air dry ton of pulp.

C. RECOVERY

1. Unit operating at 15 percent excess air at economizer outlet.

2. Particulate matter of 8 grains per SDCF leaving the economizer.

3. Efficiency of 50 percent on particulate removal in the direct contact evaporators.

4. Steam shatter jets utilized on smelt spouts.

5. Design efficiency of 97 percent for precipitator with an annual average operating efficiency of 95 percent.

D. CAUSTICIZING

1. The lime kiln scrubber has been assumed at a high collection efficiency, 99 percent on lime and 80 percent on soda.

E. POWER PLANT

1. Excess air: 30 percent for bark, 10 percent for oil and 20 percent for coal.

2. Bark burning equipment is a low set spreader stoker.

3. Reinjection from dust collector: 50 percent for bark and 0 percent for coal.

KRAFT DIAGRAM NO. 8
(Continued)

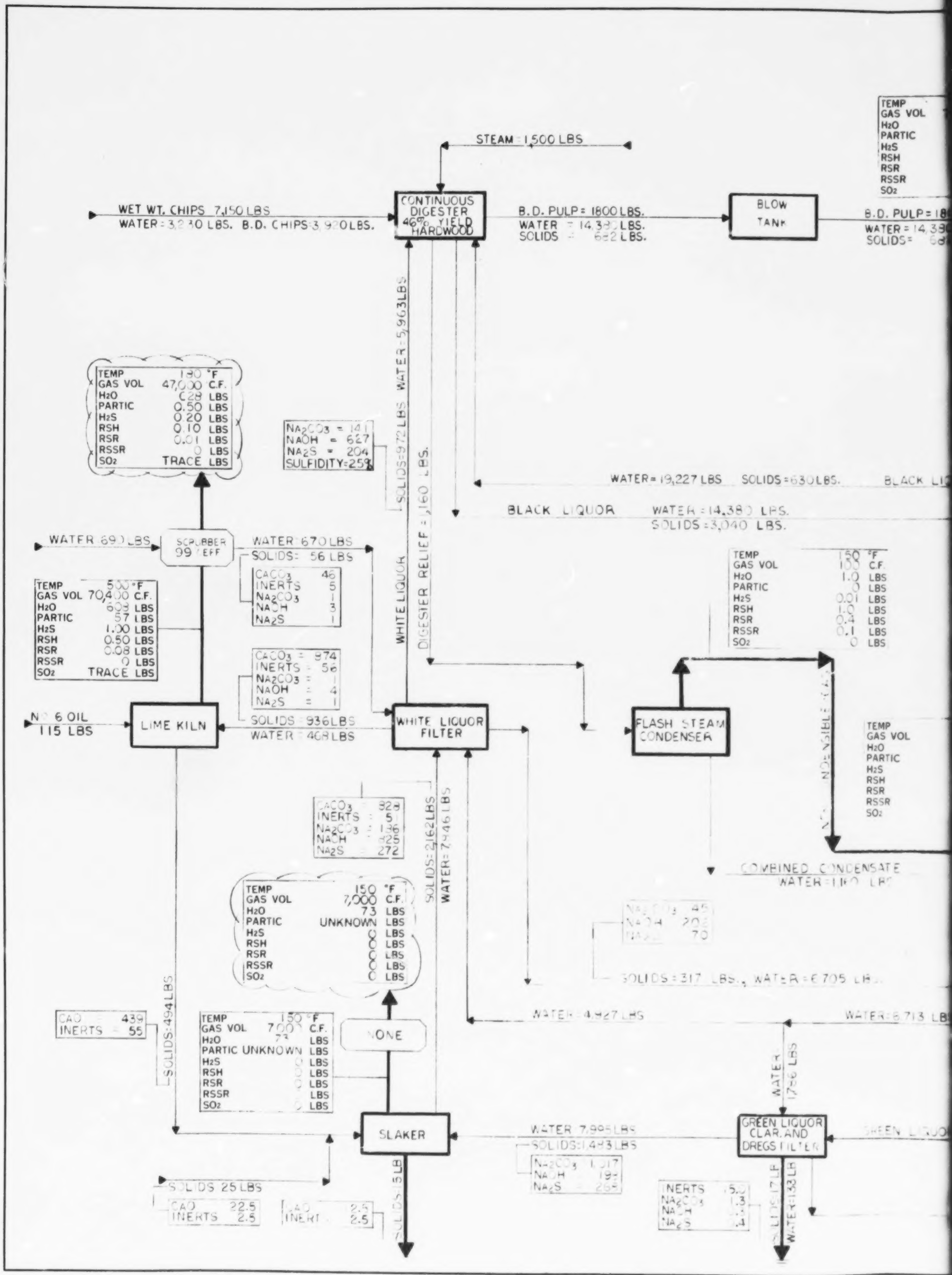
E. POWER PLANT - Continued

4. Design efficiencies for dust collector: 92 percent on bark and 95 percent on coal with annual operating efficiencies of 90 percent and 93 percent, respectively.

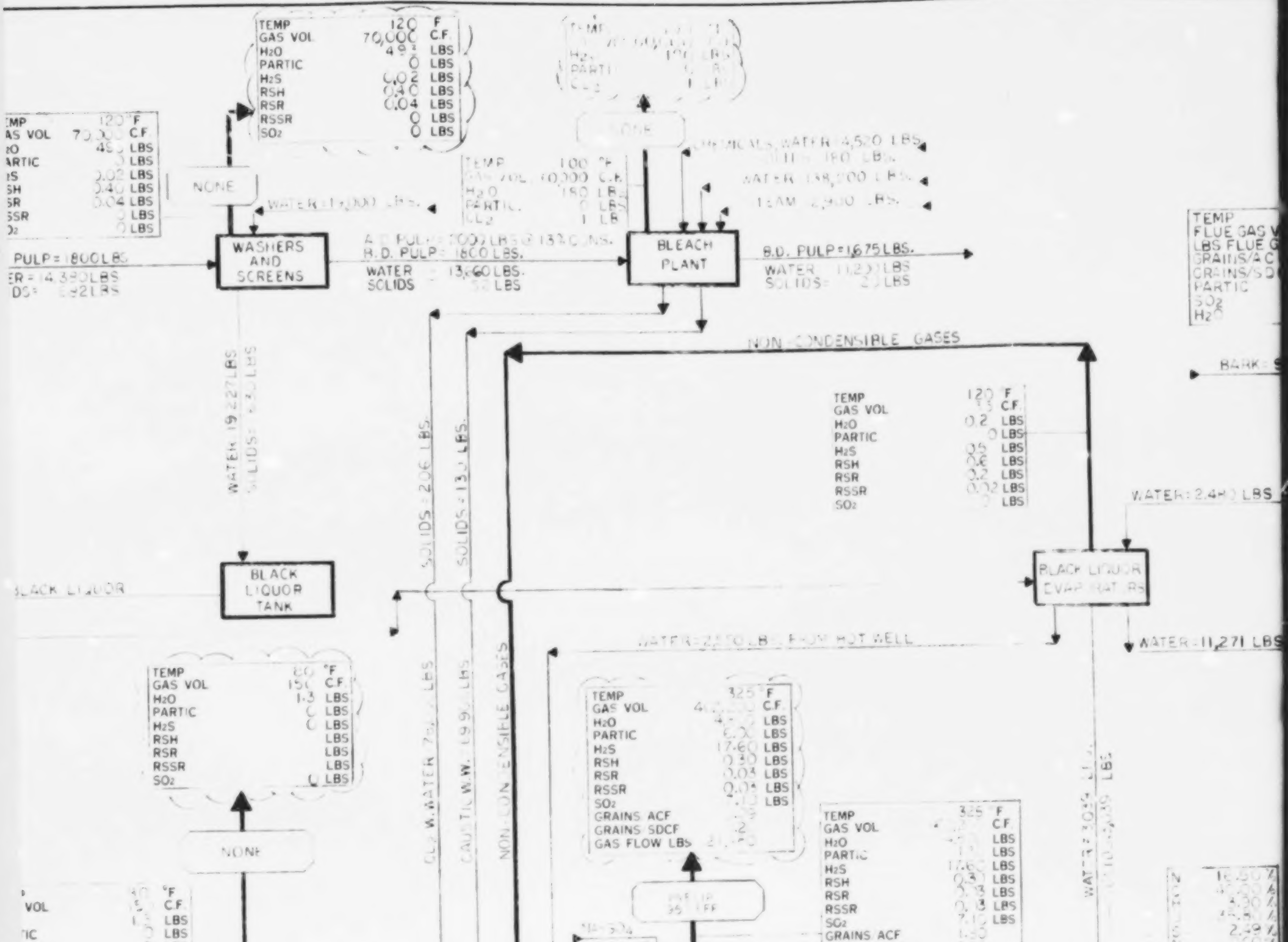
5. Bark is burned 24 hours per day at a controlled rate.

6. Unburned combustible in refuse leaving the dust collector is 40 percent.

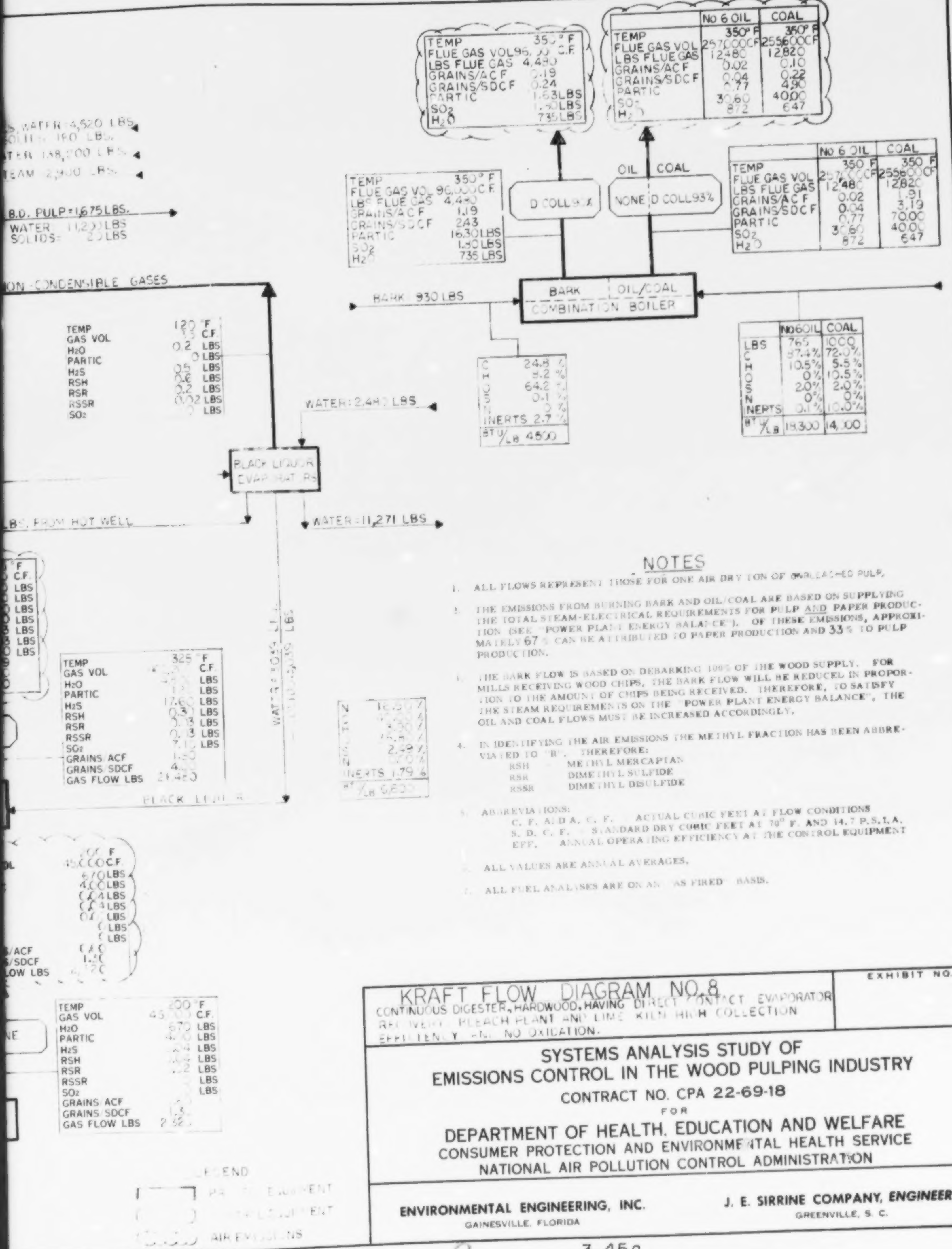
7. Coal firing based on pulverized coal.


STEAM 1,500 LBS
WET WT. CHIPS 7,150 LBS
WATER 3,230 LBS. B.D. CHIPS 3,920 LBS.
CONTINUOUS DIGESTER 46% YIELD HARDWOOD
B.D. PULP = 1800 LBS.
WATER = 14,380 LBS.
SOLIDS = 682 LBS.
BLOW TANK
B.D. PULP = 18
WATER = 14,38
SOLIDS = 68
TEMP
GAS VOL
H2O
PARTIC
H2S
RSH
RSR
RSSR
SO2
TEMP 180 °F
GAS VOL 47,000 C.F.
H2O 628 LBS
PARTIC 0.50 LBS
H2S 0.20 LBS
RSH 0.10 LBS
RSR 0.01 LBS
RSSR 0 LBS
SO2 TRACE LBS
NA2CO3 = 141
NAOH = 627
NA2S = 204
SULFIDITY = 25%
SOLIDS = 972 LBS WATER = 5,963 LBS
WHITE LIQUOR
DIGESTER RELIEF = 160 LBS.
WATER = 19,227 LBS SOLIDS = 630 LBS.
BLACK LI
BLACK LIQUOR
WATER = 14,380 LBS.
SOLIDS = 3,040 LBS.
WATER 690 LBS
SCRUBBER 99% EFF
WATER 670 LBS
SOLIDS 56 LBS
CACO3 46
INERTS 5
NA2CO3 1
NAOH 3
NA2S 1
TEMP 500 °F
GAS VOL 70,400 C.F.
H2O 608 LBS
PARTIC 57 LBS
H2S 1.00 LBS
RSH 0.50 LBS
RSR 0.08 LBS
RSSR 0 LBS
SO2 TRACE LBS
CACO3 = 874
INERTS = 56
NA2CO3 = 1
NAOH = 4
NA2S = 1
NO 6 OIL 115 LBS
LIME KILN
SOLIDS 936 LBS
WATER 468 LBS
WHITE LIQUOR FILTER
TEMP 150 °F
GAS VOL 100 C.F.
H2O 1.0 LBS
PARTIC 0 LBS
H2S 0.01 LBS
RSH 1.0 LBS
RSR 0.4 LBS
RSSR 0.1 LBS
SO2 0 LBS
FLASH STEAM CONDENSER
NON CONDENSIBLE GAS
TEMP
GAS VOL
H2O
PARTIC
H2S
RSH
RSR
RSSR
SO2
CACO3 = 828
INERTS = 51
NA2CO3 = 136
NAOH = 625
NA2S = 272
SOLIDS 2162 LBS
WATER 7446 LBS
COMBINED CONDENSATE
WATER 160 LBS
TEMP 150 °F
GAS VOL 7,000 C.F.
H2O 73 LBS
PARTIC UNKNOWN LBS
H2S 0 LBS
RSH 0 LBS
RSR 0 LBS
RSSR 0 LBS
SO2 0 LBS
NA2CO3 145
NAOH 203
NA2S 70
SOLIDS = 317 LBS., WATER = 6,705 LBS.
CAO = 439
INERTS = 55
SOLIDS 494 LBS
TEMP 150 °F
GAS VOL 7000 C.F.
H2O 73 LBS
PARTIC UNKNOWN LBS
H2S 0 LBS
RSH 0 LBS
RSR 0 LBS
RSSR 0 LBS
SO2 0 LBS
NONE
WATER 4,927 LBS
WATER 6,713 LBS
WATER 1786 LBS
SLAKER
WATER 7,995 LBS
SOLIDS 1,483 LBS
NA2CO3 1,017
NAOH 198
NA2S 268
GREEN LIQUOR CLAR. AND DREGS FILTER
GREEN LIQUOR
SOLIDS 25 LBS
CAO 22.5
INERTS 2.5
CAO 2.5
INERTS 2.5
SOLIDS 15 LB
INERTS 15.0
NA2CO3 1.3
NAOH 0.3
NA2S 0.4
SOLIDS 17 LB
WATER 133 LB

H


TEMP 120 °F
GAS VOL 70,000 C.F.
H2O 493 LBS
PARTIC 0 LBS
H2S 0.02 LBS
RSH 0.40 LBS
RSR 0.04 LBS
RSSR 0 LBS
SO2 0 LBS
NONE
WATER = 19,000 LBS.
WASHERS AND SCREENS
PULP = 1800 LBS
A.D. PULP = 2002 LBS @ 13% CONS.
B.D. PULP = 1800 LBS.
WATER 13,260 LBS
SOLIDS 52 LBS
TEMP 100 °F
GAS VOL 70,000 C.F.
H2O 180 LBS
PARTIC 0 LBS
CL2 1 LB
NONE
CHEMICALS WATER 4,520 LBS
SOLIDS 180 LBS
WATER 138,200 LBS
STEAM 2,900 LBS
BLEACH PLANT
B.D. PULP = 1,675 LBS.
WATER 11,200 LBS
SOLIDS 20 LBS
NON-CONDENSIBLE GASES
WATER 19,227 LBS
SOLIDS 430 LBS
SOLIDS = 206 LBS
SOLIDS = 130 LBS
CL2 W. WATER
CAUSTIC W.W. 19,900 LBS
NON-CONDENSIBLE GASES
TEMP 120 °F
GAS VOL 15 C.F.
H2O 0.2 LBS
PARTIC 0 LBS
H2S 0.5 LBS
RSH 0.6 LBS
RSR 0.2 LBS
RSSR 0.02 LBS
SO2 0 LBS
WATER = 2,480 LBS
BLACK LIQUOR EVAPORATORS
WATER = 11,271 LBS
BLACK LIQUOR
BLACK LIQUOR TANK
WATER = 2,470 LBS FROM HOT WELL
TEMP 80 °F
GAS VOL 150 C.F.
H2O 1.3 LBS
PARTIC 0 LBS
H2S 0 LBS
RSH LBS
RSR LBS
RSSR LBS
SO2 0 LBS
NONE
ODOR CONTROL SYSTEM
TEMP 325 °F
GAS VOL C.F.
H2O LBS
PARTIC 6.00 LBS
H2S 17.60 LBS
RSH 0.30 LBS
RSR 0.03 LBS
RSSR 0.03 LBS
SO2 7.10 LBS
GRAINS/ACF
GRAINS/SDCF
GAS FLOW LBS 21,480
TEMP 325 °F
GAS VOL C.F.
H2O LBS
PARTIC LBS
H2S 17.60 LBS
RSH 0.30 LBS
RSR 0.03 LBS
RSSR 0.03 LBS
SO2 7.10 LBS
GRAINS/ACF 1.30
GRAINS/SDCF 4.00
GAS FLOW LBS 21,480
PRECIP 95% EFF
Na2SO4 (SALT CAKE)
RECOVERY BOILER AND CONTACT EVAPORATOR
BLACK LIQUOR
WATER
TEMP 200 °F
GAS VOL 45,000 C.F.
H2O 670 LBS
PARTIC 4.00 LBS
H2S 0.04 LBS
RSH 0.04 LBS
RSR 0.02 LBS
RSSR 0 LBS
SO2 0 LBS
GRAINS/ACF
GRAINS/SDCF
GAS FLOW LBS
SMELT 93 LBS
INERTS
NONE
TEMP 200 °F
GAS VOL 45,000 C.F.
H2O 670 LBS
PARTIC 4.00 LBS
H2S 0.04 LBS
RSH 0.04 LBS
RSR 0.02 LBS
RSSR LBS
SO2 LBS
GRAINS/ACF
GRAINS/SDCF 1.3
GAS FLOW LBS 2,320
SMELT DISSOLVING TANK
GREEN LIQUOR
WATER 8,151 LBS
SOLIDS 1,521 LBS
INERTS
NA2CO3
NAOH
NA2S
SOLIDS = 21 LBS
WATER = 1,779 LBS
25,713 LBS
TEMP
FLUE GAS V
LBS FLUE G
GRAINS/AC
GRAINS/SD
PARTIC
SO2
H2O
BARK
INERTS 1.79
6,800
LEGEND
PROCESS EQUIPMENT
CONTROL EQUIPMENT
AIR EMISSIONS
KRA
CONTINUO
RECOVER
EFFLUEN
ENVIR

B


TEMP 350° F
FLUE GAS VOL 96,000 C.F.
LBS FLUE GAS 4,480
GRAINS/ACF 0.19
GRAINS/SDCF 0.24
PARTIC 1.63 LBS
SO2 1.80 LBS
H2O 735 LBS
No 6 OIL
COAL
TEMP 350° F 350° F
FLUE GAS VOL 257000CF 255600CF
LBS FLUE GAS 12480 12,820
GRAINS/ACF 0.02 0.10
GRAINS/SDCF 0.04 0.22
PARTIC 0.77 4.90
SO2 30.60 40.00
H2O 872 647
TEMP 350° F
FLUE GAS VOL 96,000 C.F.
LBS FLUE GAS 4,480
GRAINS/ACF 1.19
GRAINS/SDCF 2.43
PARTIC 16.30 LBS
SO2 1.80 LBS
H2O 735 LBS
D COLL 90%
OIL
COAL
NONE
D COLL 93%
TEMP 350 F 350 F
FLUE GAS VOL 257000CF 255600CF
LBS FLUE GAS 12,480 12820
GRAINS/ACF 0.02 1.91
GRAINS/SDCF 0.04 3.19
PARTIC 0.77 70.00
SO2 30.60 40.00
H2O 872 647
BARK
OIL/COAL
COMBINATION BOILER
BARK 930 LBS
C 24.8 %
H 8.2 %
O 64.2 %
S 0.1 %
N 0 %
INERTS 2.7 %
BTU/LB 4500
No 6 OIL
COAL
LBS 765 1000
C 87.4% 72.0%
H 10.5% 5.5%
O 0% 10.5%
S 2.0% 2.0%
N 0% 0%
INERTS 0.1% 10.0%
BTU/LB 18300 14,000
WATER 4,520 LBS
SOLIDS 160 LBS
WATER 188,200 LBS
STEAM 2,900 LBS
B.D. PULP = 1,675 LBS.
WATER 11,230 LBS
SOLIDS = 20 LBS
NON-CONDENSIBLE GASES
TEMP 120 °F
GAS VOL C.F.
H2O 0.2 LBS
PARTIC 0 LBS
H2S 0.5 LBS
RSH 0.6 LBS
RSR 0.2 LBS
RSSR 0.02 LBS
SO2 0 LBS
WATER = 2,480 LBS
BLACK LIQUOR EVAPORATORS
LBS. FROM HOT WELL
WATER = 11,271 LBS
BLACK LIQUOR
TEMP 325 °F
GAS VOL C.F.
H2O LBS
PARTIC LBS
H2S 17.60 LBS
RSH 0.30 LBS
RSR 0.03 LBS
RSSR 0.03 LBS
SO2 7.10 LBS
GRAINS/ACF 1.80
GRAINS/SDCF 4.20
GAS FLOW LBS 21,480
INERTS 1.79 %
BTU/LB 6,600
200 F
45,000 C.F.
670 LBS
4.00 LBS
0.04 LBS
0.04 LBS
0.01 LBS
0 LBS
0 LBS
TEMP 200 °F
GAS VOL 45,000 C.F.
H2O 670 LBS
PARTIC 4.00 LBS
H2S 0.04 LBS
RSH 0.04 LBS
RSR 0.02 LBS
RSSR LBS
SO2 LBS
GRAINS/ACF
GRAINS/SDCF 1.30
GAS FLOW LBS 2,320
LEGEND
PROCESS EQUIPMENT
CONTROL EQUIPMENT
AIR EMISSIONS

NOTES

1. ALL FLOWS REPRESENT THOSE FOR ONE AIR DRY TON OF UNBLEACHED PULP.
2. THE EMISSIONS FROM BURNING BARK AND OIL/COAL ARE BASED ON SUPPLYING THE TOTAL STEAM-ELECTRICAL REQUIREMENTS FOR PULP AND PAPER PRODUCTION (SEE "POWER PLANT ENERGY BALANCE"). OF THESE EMISSIONS, APPROXIMATELY 67% CAN BE ATTRIBUTED TO PAPER PRODUCTION AND 33% TO PULP PRODUCTION.
3. THE BARK FLOW IS BASED ON DEBARKING 100% OF THE WOOD SUPPLY. FOR MILLS RECEIVING WOOD CHIPS, THE BARK FLOW WILL BE REDUCED IN PROPORTION TO THE AMOUNT OF CHIPS BEING RECEIVED. THEREFORE, TO SATISFY THE STEAM REQUIREMENTS ON THE "POWER PLANT ENERGY BALANCE", THE OIL AND COAL FLOWS MUST BE INCREASED ACCORDINGLY.
4. IN IDENTIFYING THE AIR EMISSIONS THE METHYL FRACTION HAS BEEN ABBREVIATED TO "R". THEREFORE:
   - RSH — METHYL MERCAPTAN
   - RSR — DIMETHYL SULFIDE
   - RSSR — DIMETHYL DISULFIDE
5. ABBREVIATIONS:
   - C. F. AND A. C. F. — ACTUAL CUBIC FEET AT FLOW CONDITIONS
   - S. D. C. F. — STANDARD DRY CUBIC FEET AT 70° F. AND 14.7 P.S.I.A.
   - EFF. — ANNUAL OPERATING EFFICIENCY AT THE CONTROL EQUIPMENT
6. ALL VALUES ARE ANNUAL AVERAGES.
7. ALL FUEL ANALYSES ARE ON AN "AS FIRED" BASIS.

KRAFT FLOW DIAGRAM NO. 8

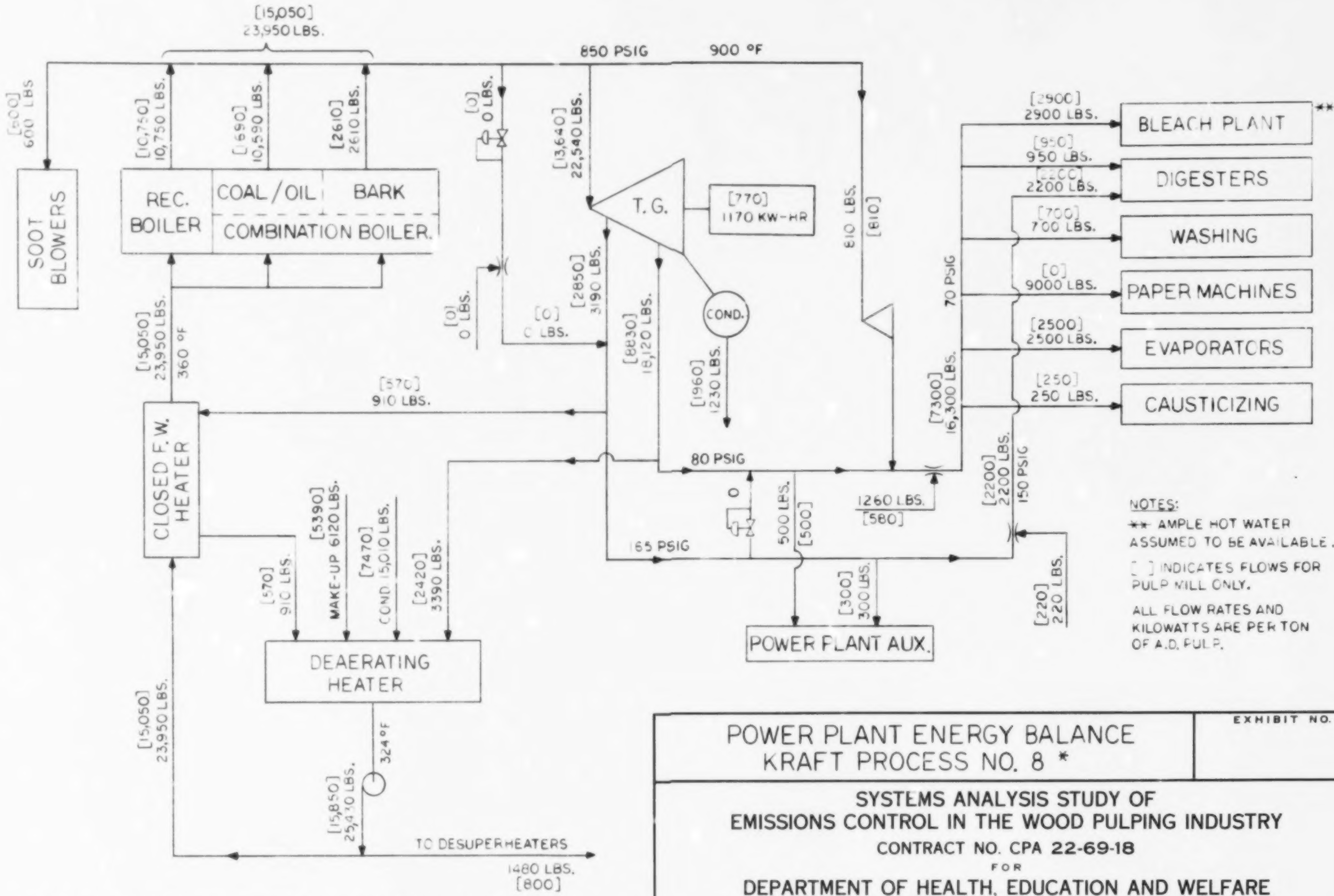
CONTINUOUS DIGESTER, HARDWOOD, HAVING DIRECT CONTACT EVAPORATOR RECOVERY, BLEACH PLANT AND LIME KILN HIGH COLLECTION EFFICIENCY AND NO OXIDATION.

EXHIBIT NO.

SYSTEMS ANALYSIS STUDY OF
EMISSIONS CONTROL IN THE WOOD PULPING INDUSTRY
CONTRACT NO. CPA 22-69-18
FOR
DEPARTMENT OF HEALTH, EDUCATION AND WELFARE
CONSUMER PROTECTION AND ENVIRONMENTAL HEALTH SERVICE
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

ENVIRONMENTAL ENGINEERING, INC.
GAINESVILLE, FLORIDA

J. E. SIRRINE COMPANY, ENGINEERS
GREENVILLE, S. C.


[15,050]
23,950 LBS.
850 PSIG
900 °F
[600]
600 LBS.
SOOT BLOWERS
[10,750]
10,750 LBS.
[1690]
10,590 LBS.
[2610]
2610 LBS.
REC. BOILER
COAL / OIL
BARK
COMBINATION BOILER.
[0]
0 LBS.
[13,640]
22,540 LBS.
T. G.
[770]
1170 KW-HR
810 LBS.
[810]
[0]
0 LBS.
[0]
0 LBS.
[2850]
3190 LBS.
[8830]
18,120 LBS.
COND.
[1960]
1230 LBS.
[15,050]
23,950 LBS.
360 °F
[570]
910 LBS.
CLOSED F.W. HEATER
80 PSIG
70 PSIG
[7300]
16,300 LBS.
1260 LBS.
[580]
0
500 LBS.
[500]
165 PSIG
[300]
300 LBS.
POWER PLANT AUX.
[2200]
2200 LBS.
150 PSIG
[220]
220 LBS.
[570]
910 LBS.
[5390]
MAKE-UP 6120 LBS.
[7470]
COND. 15,010 LBS.
[2420]
3390 LBS.
DEAERATING HEATER
324 °F
[15,850]
25,430 LBS.
[15,050]
23,950 LBS.
TO DESUPERHEATERS
1480 LBS.
[800]
[2900]
2900 LBS.
BLEACH PLANT **
[950]
950 LBS.
DIGESTERS
[2200]
2200 LBS.
[700]
700 LBS.
WASHING
[0]
9000 LBS.
PAPER MACHINES
[2500]
2500 LBS.
EVAPORATORS
[250]
250 LBS.
CAUSTICIZING

NOTES:

** AMPLE HOT WATER ASSUMED TO BE AVAILABLE.

[ ] INDICATES FLOWS FOR PULP MILL ONLY.

ALL FLOW RATES AND KILOWATTS ARE PER TON OF A.D. PULP.

* CONTINUOUS DIGESTERS, HARDWOOD, NO LIQUOR OXIDATION, DIRECT CONTACT EVAPORATOR RECOVERY, BLEACHING, AND LIME KILN HIGH COLLECTION EFFICIENCY.

POWER PLANT ENERGY BALANCE
KRAFT PROCESS NO. 8 *

EXHIBIT NO.

SYSTEMS ANALYSIS STUDY OF
EMISSIONS CONTROL IN THE WOOD PULPING INDUSTRY
CONTRACT NO. CPA 22-69-18
FOR
DEPARTMENT OF HEALTH, EDUCATION AND WELFARE
CONSUMER PROTECTION AND ENVIRONMENTAL HEALTH SERVICE
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

ENVIRONMENTAL ENGINEERING, INC.
GAINESVILLE, FLORIDA

J. E. SIRRINE COMPANY, *ENGINEERS*
GREENVILLE, S. C.

KRAFT DIAGRAM NO. 9

TYPICAL LOCATION

Midwestern United States.

TYPICAL AGE OF EQUIPMENT

Under ten years.

GENERAL

This flow diagram uses equipment similar to Flow Diagram No. 8 which produces hardwood pulp, whereas No. 9 produces softwood pulp. We have used similar flow diagrams, except for wood species, to show the increase of emissions when cooking hardwood. There are perhaps 30 to 40 mills in the U. S. that are illustrated by this flow diagram.

EMISSIONS

The following emissions have been selected as representative, based on the range of emissions presented in Chapter 4.

POUNDS PER AIR DRY TON OF UNBLEACHED PULP

LOCATION	H_2S	RSH, RSR, RSSR	SO_2	Particulate
Washers	0.02	0.22	0	0
M. E. Evaporators	0.50	0.39	0	0
R. Boiler and D. C. Evaporator	17.1	0.24	7.90	114
After Precipitator	17.1	0.24	7.90	5.70
Smelt tank	0.02	0.02	0	1.0
Slaker	0	0	0	Unknown
Lime Kiln	1.0	0.6	Trace	55
Lime Kiln After Scrubber	0.2	0.11	Trace	0.50
Turpentine Condenser	0.01	0.35	0	0

KRAFT DIAGRAM NO. 9
(Continued)

ASSUMPTIONS

The following assumptions have been made in developing the flow diagram:

A. PULP MILL

1. Pulp Yield = 45 Percent.
2. Cooking liquor charge per air dry ton of brown stock equals 6,996 pounds of which 996 pounds are chemical solids. Sulfidity equals 25 percent.

B. BLEACH PLANT

Similar to Flow Diagram No. 8.

C. RECOVERY

1. Unit operating at 15 percent excess air at economizer outlet.
2. Particulate matter of 8 grains per SDCF leaving the economizer.
3. Efficiency of 50 percent on particulate removal in the direct contact evaporators.
4. Steam shatter jets utilized on smelt spouts.
5. Design efficiency of 97 percent for precipitator with an annual average operating efficiency of 95 percent.

D. CAUSTICIZING

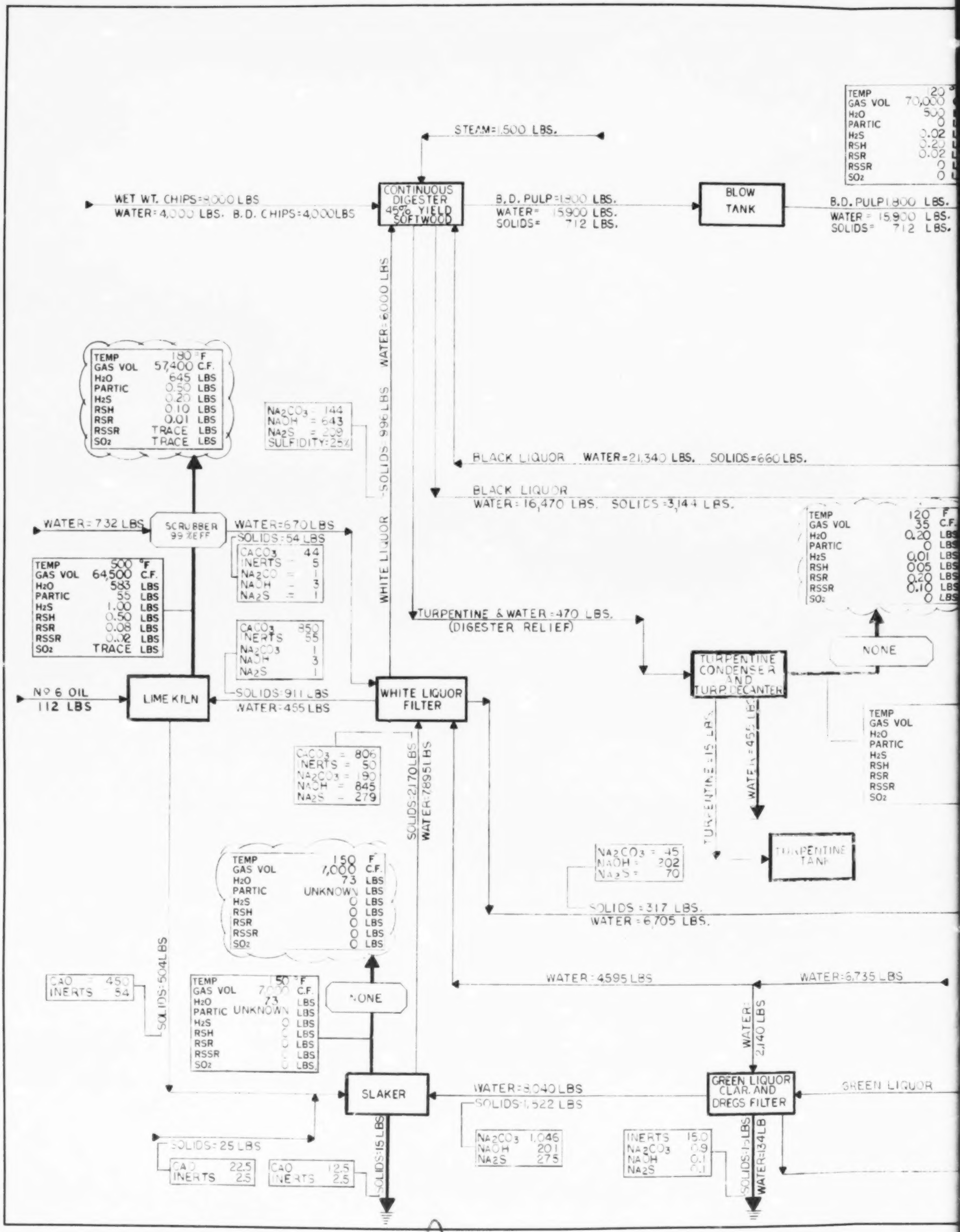
Similar to Flow Diagram No. 8

E. POWER PLANT

1. Excess air: 30 percent for bark, 10 percent for oil, and 20 percent for coal.
2. Bark burning equipment is a low set spreader stoker.

KRAFT DIAGRAM NO. 9
(Continued)

3. Reinjection from dust collector: 50 percent for bark and 0 percent for coal.
4. Design efficiencies for dust collector: 92 percent on bark and 95 percent on coal with annual operating efficiencies of 90 percent and 93 percent, respectively.
5. Bark is burned 24 hours per day at a controlled rate.
6. Unburned combustible in refuse leaving the dust collector is 40 percent.
7. Coal firing based on pulverized coal.


STEAM=1,500 LBS.
WET WT. CHIPS=8,000 LBS
WATER=4,000 LBS. B.D. CHIPS=4,000LBS
CONTINUOUS DIGESTER 45% YIELD SOFTWOOD
B.D. PULP=1,800 LBS.
WATER= 15900 LBS.
SOLIDS= 712 LBS.
BLOW TANK
B.D. PULP1,800 LBS.
WATER= 15,900 LBS.
SOLIDS= 712 LBS.
TEMP 120
GAS VOL 70,000
H2O 500
PARTIC 0
H2S 0.02
RSH 0.20
RSR 0.02
RSSR 0
SO2 0
WATER=6000 LBS
SOLIDS 996 LBS
WHITE LIQUOR
NA2CO3 = 144
NAOH = 643
NA2S = 209
SULFIDITY:25%
TEMP 180 °F
GAS VOL 57,400 C.F.
H2O 645 LBS
PARTIC 0.50 LBS
H2S 0.20 LBS
RSH 0.10 LBS
RSR 0.01 LBS
RSSR TRACE LBS
SO2 TRACE LBS
BLACK LIQUOR WATER=21,340 LBS. SOLIDS=660 LBS.
BLACK LIQUOR
WATER=16,470 LBS. SOLIDS=3,144 LBS.
WATER=732 LBS
SCRUBBER 99% EFF
WATER=670 LBS
SOLIDS:54 LBS
CACO3 44
INERTS 5
NA2CO 1
NAOH 3
NA2S 1
TEMP 500 °F
GAS VOL 64,500 C.F.
H2O 583 LBS
PARTIC 55 LBS
H2S 1.00 LBS
RSH 0.50 LBS
RSR 0.08 LBS
RSSR 0.02 LBS
SO2 TRACE LBS
CACO3 850
INERTS 55
NA2CO3 1
NAOH 3
NA2S 1
TURPENTINE & WATER=470 LBS.
(DIGESTER RELIEF)
TEMP 120 F
GAS VOL 35 C.F.
H2O 0.20 LBS
PARTIC 0 LBS
H2S 0.01 LBS
RSH 0.05 LBS
RSR 0.20 LBS
RSSR 0.10 LBS
SO2 0 LBS
NONE
TURPENTINE CONDENSER AND TURP. DECANTER
Nº 6 OIL
112 LBS
LIME KILN
SOLIDS:911 LBS
WATER:455 LBS
WHITE LIQUOR FILTER
TEMP
GAS VOL
H2O
PARTIC
H2S
RSH
RSR
RSSR
SO2
CACO3 = 806
INERTS = 50
NA2CO3 = 190
NAOH = 845
NA2S = 279
SOLIDS:2170 LBS
WATER:7895 LBS
TURPENTINE:15 LBS
WATER=455 LBS
TURPENTINE TANK
NA2CO3 = 45
NAOH = 202
NA2S = 70
SOLIDS = 317 LBS.
WATER = 6,705 LBS.
TEMP 150 F
GAS VOL 7,000 C.F.
H2O 73 LBS
PARTIC UNKNOWN LBS
H2S 0 LBS
RSH 0 LBS
RSR 0 LBS
RSSR 0 LBS
SO2 0 LBS
CAO = 450
INERTS = 54
SOLIDS:504LBS
TEMP 150 °F
GAS VOL 7,000 C.F.
H2O 73 LBS
PARTIC UNKNOWN LBS
H2S 0 LBS
RSH 0 LBS
RSR 0 LBS
RSSR 0 LBS
SO2 0 LBS
NONE
WATER:4595 LBS
WATER:6,735 LBS
WATER: 2,140 LBS
SLAKER
WATER=8,040 LBS
SOLIDS:1,522 LBS
GREEN LIQUOR CLAR. AND DREGS FILTER
GREEN LIQUOR
NA2CO3 1,046
NAOH 201
NA2S 275
INERTS 15.0
NA2CO3 0.9
NAOH 0.1
NA2S 0.1
SOLIDS:15 LBS
SOLIDS:1-5 LBS
WATER:134 LB
SOLIDS=25 LBS
CAO 22.5
INERTS 2.5
CAO 12.5
INERTS 2.5

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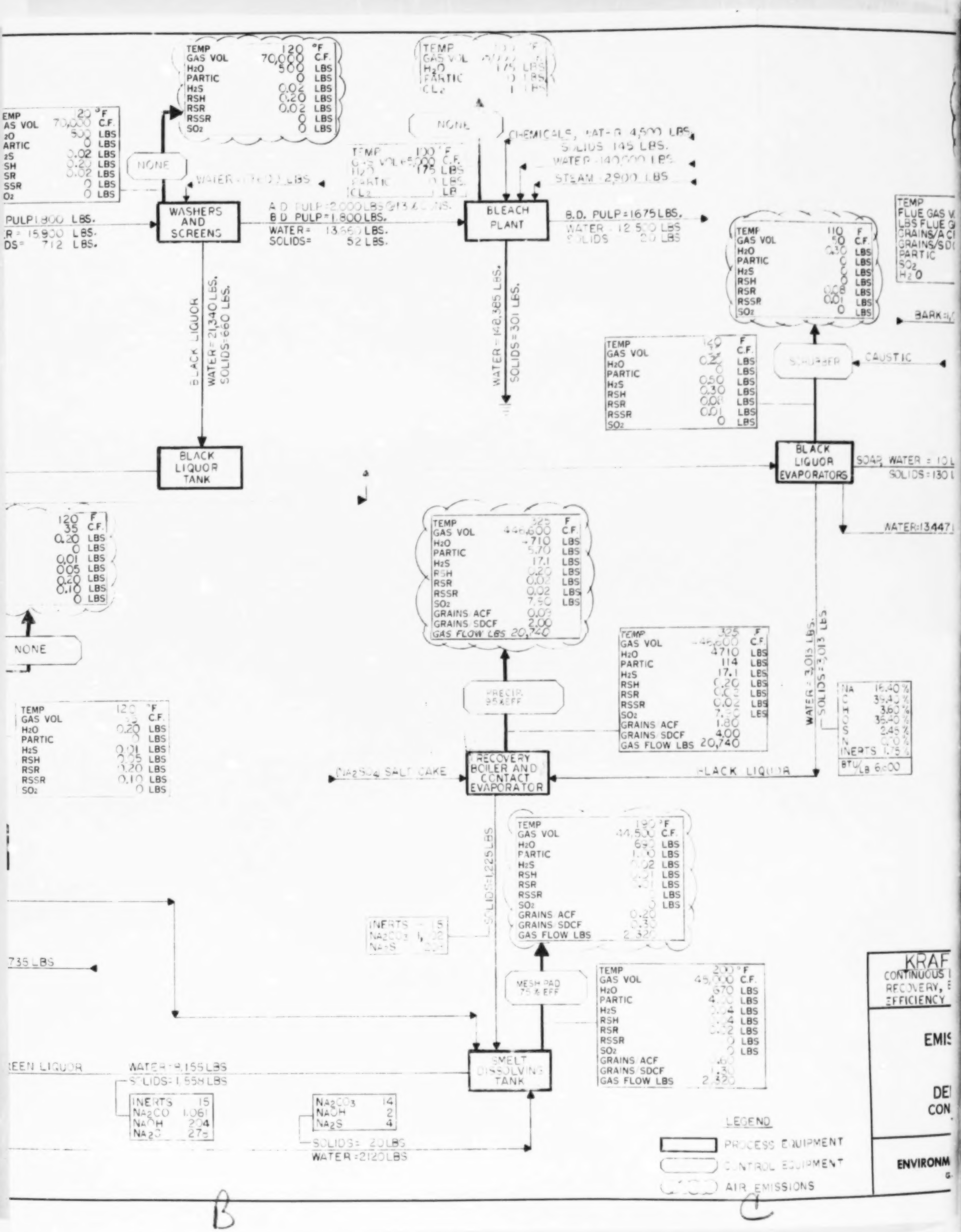
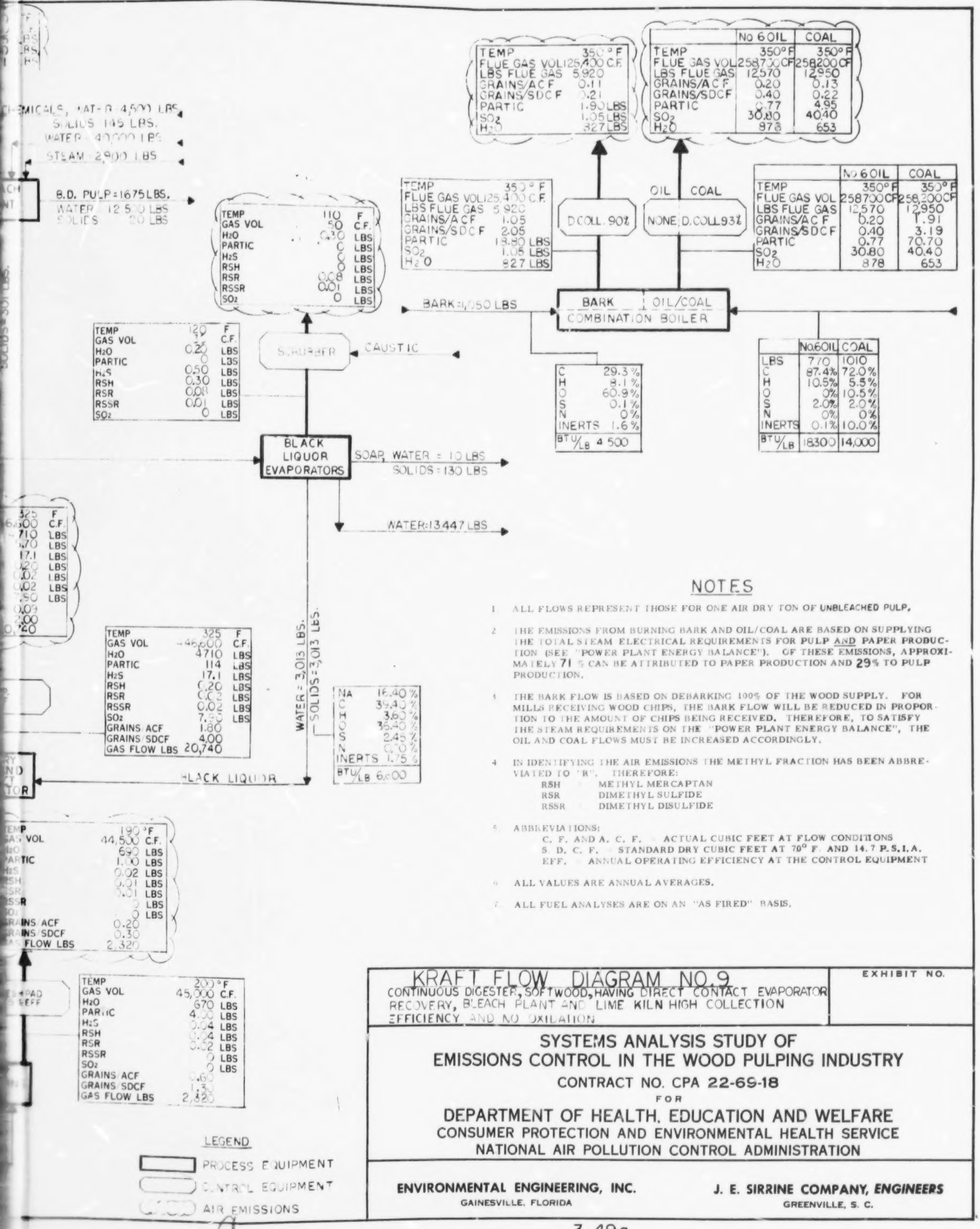

TEMP 120 °F
GAS VOL 70,000 C.F.
H2O 500 LBS
PARTIC 0 LBS
H2S 0.02 LBS
RSH 0.20 LBS
RSR 0.02 LBS
RSSR 0 LBS
SO2 0 LBS
NONE
WATER = 17,600 LBS
WASHERS AND SCREENS
PULP 1,800 LBS.
R = 15,900 LBS.
DS = 712 LBS.
A.D. PULP = 2,000 LBS @ 13.4 CONS.
B.D. PULP = 1,800 LBS.
WATER = 13,660 LBS.
SOLIDS = 52 LBS.
BLACK LIQUOR
WATER = 21,340 LBS.
SOLIDS = 660 LBS.
BLACK LIQUOR TANK
TEMP 100 °F
GAS VOL 65,000 C.F.
H2O 175 LBS
PARTIC 0 LBS
CL2 1 LB
NONE
CHEMICALS, MAT-R 4,500 LBS
SOLIDS 145 LBS.
WATER = 140,000 LBS.
STEAM = 2,900 LBS
BLEACH PLANT
B.D. PULP = 1675 LBS.
WATER = 12,500 LBS
SOLIDS 20 LBS
WATER = 148,385 LBS.
SOLIDS = 301 LBS.
TEMP 110 °F
GAS VOL 50 C.F.
H2O 0.30 LBS
PARTIC 0 LBS
H2S 0 LBS
RSH 0 LBS
RSR 0.08 LBS
RSSR 0.01 LBS
SO2 0 LBS
TEMP 140 °F
GAS VOL 0.22 C.F.
H2O 0 LBS
PARTIC 0 LBS
H2S 0.50 LBS
RSH 0.30 LBS
RSR 0.08 LBS
RSSR 0.01 LBS
SO2 0 LBS
SCRUBBER
CAUSTIC
BLACK LIQUOR EVAPORATORS
SOAP, WATER = 10
SOLIDS = 130
WATER = 13447
TEMP
FLUE GAS V
LBS FLUE G
GRAINS/A C
GRAINS/SD
PARTIC
SO2
H2O
BARK = 1
TEMP 120 °F
GAS VOL 35 C.F.
H2O 0.20 LBS
PARTIC 0 LBS
H2S 0.01 LBS
RSH 0.05 LBS
RSR 0.20 LBS
RSSR 0.10 LBS
SO2 0 LBS
NONE
TEMP 120 °F
GAS VOL 35 C.F.
H2O 0.20 LBS
PARTIC 0 LBS
H2S 0.01 LBS
RSH 0.05 LBS
RSR 0.20 LBS
RSSR 0.10 LBS
SO2 0 LBS
TEMP 325 °F
GAS VOL 446,600 C.F.
H2O 4710 LBS
PARTIC 5.70 LBS
H2S 17.1 LBS
RSH 0.20 LBS
RSR 0.02 LBS
RSSR 0.02 LBS
SO2 7.50 LBS
GRAINS/ACF 0.09
GRAINS/SDCF 2.00
GAS FLOW LBS 20,740
PRECIP. 95% EFF
TEMP 325 °F
GAS VOL 446,600 C.F.
H2O 4710 LBS
PARTIC 114 LBS
H2S 17.1 LBS
RSH 0.20 LBS
RSR 0.02 LBS
RSSR 0.02 LBS
SO2 7.50 LBS
GRAINS/ACF 1.80
GRAINS/SDCF 4.00
GAS FLOW LBS 20,740
WATER = 3,013 LBS.
SOLIDS = 3,013 LBS.
NA 16.40 %
C 39.40 %
H 3.60 %
O 35.40 %
S 2.45 %
N 0.00 %
INERTS 1.75 %
BTU/LB 6,600
NA2SO4 SALT CAKE
RECOVERY BOILER AND CONTACT EVAPORATOR
BLACK LIQUOR
SOLIDS = 1,225 LBS
INERTS 15
NA2CO3 1,002
NA2S 201
TEMP 190 °F
GAS VOL 44,500 C.F.
H2O 690 LBS
PARTIC 1.00 LBS
H2S 0.02 LBS
RSH 0.01 LBS
RSR 0.01 LBS
RSSR 0 LBS
SO2 0 LBS
GRAINS/ACF 0.20
GRAINS/SDCF 0.30
GAS FLOW LBS 2,320
MESH PAD 75% EFF
TEMP 200 °F
GAS VOL 45,000 C.F.
H2O 670 LBS
PARTIC 4.00 LBS
H2S 0.04 LBS
RSH 0.04 LBS
RSR 0.02 LBS
RSSR 0 LBS
SO2 0 LBS
GRAINS/ACF 0.60
GRAINS/SDCF 1.30
GAS FLOW LBS 2,320
735 LBS
SMELT DISSOLVING TANK
GREEN LIQUOR
WATER = 8,155 LBS
SOLIDS = 1,558 LBS
INERTS 15
NA2CO 1,061
NAOH 204
NA2S 278
NA2CO3 14
NAOH 2
NA2S 4
SOLIDS = 20 LBS
WATER = 2120 LBS
LEGEND
PROCESS EQUIPMENT
CONTROL EQUIPMENT
AIR EMISSIONS
KRAF
CONTINUOUS
RECOVERY, E
EFFICIENCY
EMIS
DE
CON
ENVIRONM
G.

B

C


CHEMICALS, WATER 4,500 LBS
SOLIDS 145 LBS.
WATER 40,000 LBS
STEAM 2,900 LBS
B.D. PULP=1675 LBS.
WATER 12,500 LBS
SOLIDS 20 LBS
TEMP 110 F; GAS VOL 50 C.F.; H2O 0.30 LBS; PARTIC 0 LBS; H2S 0 LBS; RSH 0 LBS; RSR 0.08 LBS; RSSR 0.01 LBS; SO2 0 LBS
SCRUBBER
CAUSTIC
TEMP 120 F; GAS VOL 1? C.F.; H2O 0.25 LBS; PARTIC 0 LBS; H2S 0.50 LBS; RSH 0.30 LBS; RSR 0.08 LBS; RSSR 0.01 LBS; SO2 0 LBS
BLACK LIQUOR EVAPORATORS
SOAP, WATER = 10 LBS
SOLIDS = 130 LBS
WATER: 13447 LBS
WATER = 3,013 LBS.
SOLIDS = 3,013 LBS.
NA 16.40 %; C 39.40 %; H 3.60 %; O 35.40 %; S 2.45 %; N 0.00 %; INERTS 1.75 %; BTU/LB 6,600
BLACK LIQUOR
TEMP 325 F; GAS VOL 46,600 C.F.; H2O 4710 LBS; PARTIC 114 LBS; H2S 17.1 LBS; RSH 0.20 LBS; RSR 0.02 LBS; RSSR 0.02 LBS; SO2 7.90 LBS; GRAINS/ACF 1.80; GRAINS/SDCF 4.00; GAS FLOW LBS 20,740
325 F; 46,600 C.F.; 4710 LBS; 5.70 LBS; 17.1 LBS; 0.20 LBS; 0.02 LBS; 0.02 LBS; 7.90 LBS; 0.09; 2.00; 20,740
TEMP 190 °F; GAS VOL 44,500 C.F.; H2O 690 LBS; PARTIC 1.00 LBS; H2S 0.02 LBS; RSH 0.01 LBS; RSR 0.01 LBS; RSSR 0 LBS; SO2 0 LBS; GRAINS/ACF 0.20; GRAINS/SDCF 0.30; GAS FLOW LBS 2,320
TEMP 200 °F; GAS VOL 45,000 C.F.; H2O 670 LBS; PARTIC 4.00 LBS; H2S 0.04 LBS; RSH 0.04 LBS; RSR 0.02 LBS; RSSR 0 LBS; SO2 0 LBS; GRAINS/ACF 0.60; GRAINS/SDCF 1.30; GAS FLOW LBS 2,320
BARK = 1,050 LBS
BARK | OIL/COAL COMBINATION BOILER
C 29.3 %; H 8.1 %; O 60.9 %; S 0.1 %; N 0 %; INERTS 1.6 %; BTU/LB 4 500
No. 6 OIL / COAL: LBS 770 / 1010; C 87.4% / 72.0%; H 10.5% / 5.5%; O 0% / 10.5%; S 2.0% / 2.0%; N 0% / 0%; INERTS 0.1% / 10.0%; BTU/LB 18300 / 14,000
TEMP 350 °F; FLUE GAS VOL 125,400 C.F.; LBS FLUE GAS 5,920; GRAINS/ACF 1.05; GRAINS/SDCF 2.05; PARTIC 18.80 LBS; SO2 1.05 LBS; H2O 827 LBS
D. COLL. 90%
OIL | COAL
NONE | D. COLL 93%
No 6 OIL / COAL: TEMP 350°F / 350°F; FLUE GAS VOL 258700 CF / 258,200 CF; LBS FLUE GAS 12,570 / 12,950; GRAINS/ACF 0.20 / 1.91; GRAINS/SDCF 0.40 / 3.19; PARTIC 0.77 / 70.70; SO2 30.80 / 40.40; H2O 878 / 653
TEMP 350 °F; FLUE GAS VOL 125,400 C.F.; LBS FLUE GAS 5,920; GRAINS/ACF 0.11; GRAINS/SDCF 0.21; PARTIC 1.90 LBS; SO2 1.05 LBS; H2O 827 LBS
No 6 OIL / COAL: TEMP 350°F / 350°F; FLUE GAS VOL 258700 CF / 258200 CF; LBS FLUE GAS 12,570 / 12,950; GRAINS/ACF 0.20 / 0.13; GRAINS/SDCF 0.40 / 0.22; PARTIC 0.77 / 4.95; SO2 30.80 / 40.40; H2O 878 / 653
LEGEND
PROCESS EQUIPMENT
CONTROL EQUIPMENT
AIR EMISSIONS

NOTES

1. ALL FLOWS REPRESENT THOSE FOR ONE AIR DRY TON OF UNBLEACHED PULP.
2. THE EMISSIONS FROM BURNING BARK AND OIL/COAL ARE BASED ON SUPPLYING THE TOTAL STEAM ELECTRICAL REQUIREMENTS FOR PULP AND PAPER PRODUCTION (SEE "POWER PLANT ENERGY BALANCE"). OF THESE EMISSIONS, APPROXIMATELY 71 % CAN BE ATTRIBUTED TO PAPER PRODUCTION AND 29% TO PULP PRODUCTION.
3. THE BARK FLOW IS BASED ON DEBARKING 100% OF THE WOOD SUPPLY. FOR MILLS RECEIVING WOOD CHIPS, THE BARK FLOW WILL BE REDUCED IN PROPORTION TO THE AMOUNT OF CHIPS BEING RECEIVED. THEREFORE, TO SATISFY THE STEAM REQUIREMENTS ON THE "POWER PLANT ENERGY BALANCE", THE OIL AND COAL FLOWS MUST BE INCREASED ACCORDINGLY.
4. IN IDENTIFYING THE AIR EMISSIONS THE METHYL FRACTION HAS BEEN ABBREVIATED TO "R". THEREFORE:
   - RSH = METHYL MERCAPTAN
   - RSR = DIMETHYL SULFIDE
   - RSSR = DIMETHYL DISULFIDE
5. ABBREVIATIONS:
   - C. F. AND A. C. F. = ACTUAL CUBIC FEET AT FLOW CONDITIONS
   - S. D. C. F. = STANDARD DRY CUBIC FEET AT 70° F. AND 14.7 P.S.I.A.
   - EFF. = ANNUAL OPERATING EFFICIENCY AT THE CONTROL EQUIPMENT
6. ALL VALUES ARE ANNUAL AVERAGES.
7. ALL FUEL ANALYSES ARE ON AN "AS FIRED" BASIS.

KRAFT FLOW DIAGRAM NO. 9

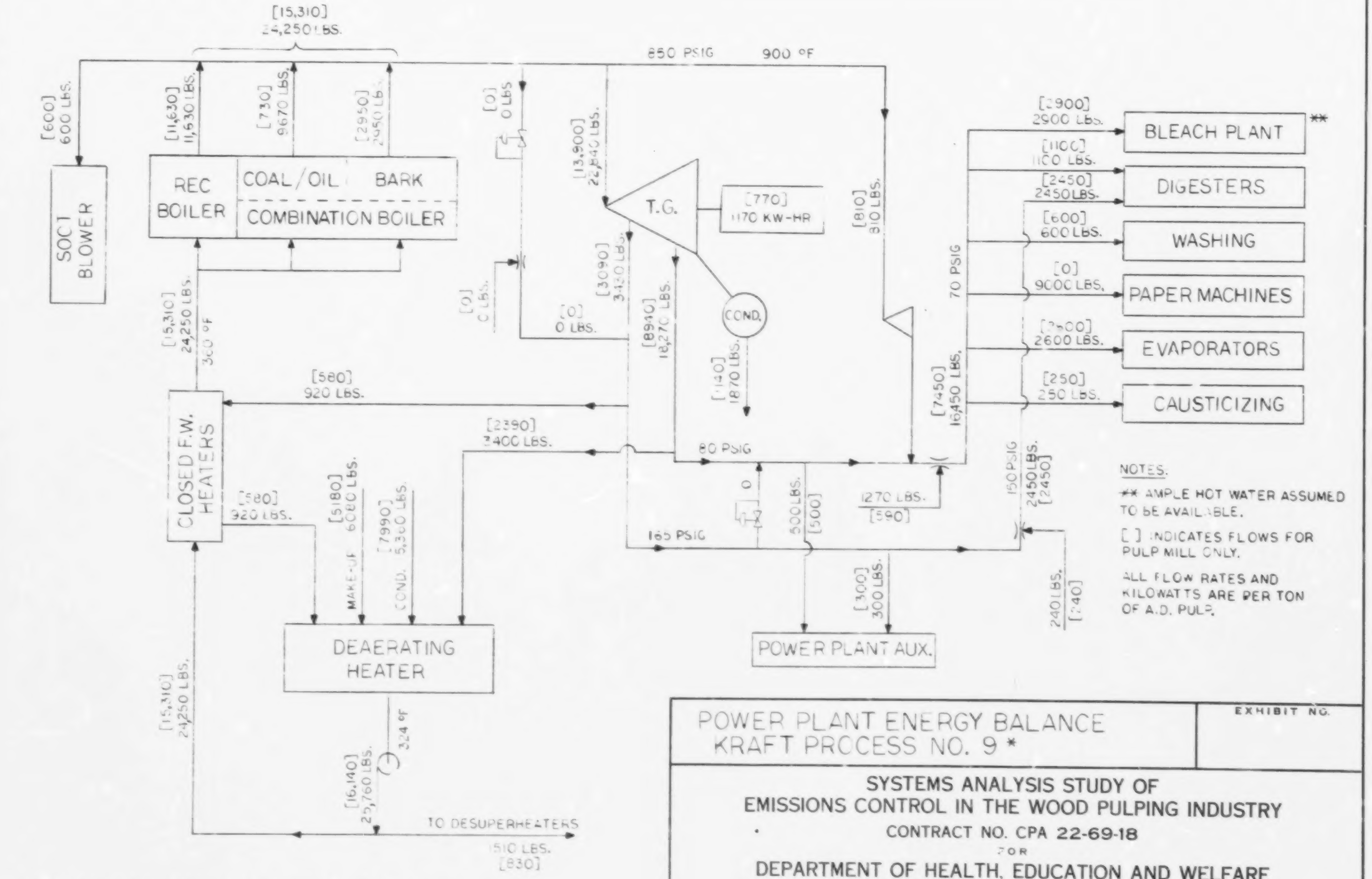
CONTINUOUS DIGESTER, SOFTWOOD, HAVING DIRECT CONTACT EVAPORATOR RECOVERY, BLEACH PLANT AND LIME KILN HIGH COLLECTION EFFICIENCY AND NO OXIDATION

EXHIBIT NO.

SYSTEMS ANALYSIS STUDY OF
EMISSIONS CONTROL IN THE WOOD PULPING INDUSTRY
CONTRACT NO. CPA 22-69-18
FOR
DEPARTMENT OF HEALTH, EDUCATION AND WELFARE
CONSUMER PROTECTION AND ENVIRONMENTAL HEALTH SERVICE
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

ENVIRONMENTAL ENGINEERING, INC.
GAINESVILLE, FLORIDA

J. E. SIRRINE COMPANY, ENGINEERS
GREENVILLE, S. C.


[15,310]
24,250 LBS.
850 PSIG
900 °F
[600]
600 LBS.
SOOT BLOWER
[11,630]
11,630 LBS.
[730]
9670 LBS.
[2950]
2950 LBS.
REC BOILER
COAL/OIL
BARK
COMBINATION BOILER
[0]
0 LBS.
[13,900]
22,840 LBS.
T.G.
[770]
1170 KW-HR
[810]
810 LBS.
[0]
0 LBS.
[3090]
3420 LBS.
[0]
0 LBS.
[8940]
18,270 LBS.
COND.
[1140]
1870 LBS.
[7450]
16,450 LBS.
70 PSIG
[15,310]
24,250 LBS.
360 °F
[580]
920 LBS.
CLOSED F.W. HEATERS
[2390]
3400 LBS.
80 PSIG
0
500 LBS.
[500]
1270 LBS.
[590]
165 PSIG
150 PSIG
2450 LBS.
[2450]
[580]
920 LBS.
[5180]
MAKE-UP 6080 LBS.
[7990]
COND. 5,360 LBS.
DEAERATING HEATER
[300]
300 LBS.
POWER PLANT AUX.
240 LBS.
[240]
[15,310]
24,250 LBS.
[16,140]
25,760 LBS.
324 °F
TO DESUPERHEATERS
1510 LBS.
[830]
[2900]
2900 LBS.
BLEACH PLANT **
[1100]
1100 LBS.
DIGESTERS
[2450]
2450 LBS.
WASHING
[600]
600 LBS.
PAPER MACHINES
[0]
9000 LBS.
EVAPORATORS
[2600]
2600 LBS.
CAUSTICIZING
[250]
250 LBS.

NOTES:

** AMPLE HOT WATER ASSUMED TO BE AVAILABLE.

[ ] INDICATES FLOWS FOR PULP MILL ONLY.

ALL FLOW RATES AND KILOWATTS ARE PER TON OF A.D. PULP.

* CONTINUOUS DIGESTERS, SOFTWOOD, NO LIQUOR OXIDATION, DIRECT CONTACT EVAPORATOR RECOVERY, BLEACHING, AND LIME KILN HIGH COLLECTION EFFICIENCY.

POWER PLANT ENERGY BALANCE
KRAFT PROCESS NO. 9 *

EXHIBIT NO.

SYSTEMS ANALYSIS STUDY OF
EMISSIONS CONTROL IN THE WOOD PULPING INDUSTRY
CONTRACT NO. CPA 22-69-18
FOR
DEPARTMENT OF HEALTH, EDUCATION AND WELFARE
CONSUMER PROTECTION AND ENVIRONMENTAL HEALTH SERVICE
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

ENVIRONMENTAL ENGINEERING, INC.
GAINESVILLE, FLORIDA

J. E. SIRRINE COMPANY, ENGINEERS
GREENVILLE, S. C.

KRAFT DIAGRAM NO. 10

TYPICAL LOCATION

Southern United States

TYPICAL AGE OF EQUIPMENT

1969 design; 1972 start up

GENERAL

This flow diagram is based on production of unbleached softwood kraft pulp and depicts a new mill designed with emphasis on odor abatement and reduced particulate emission. It includes the installation of high solids multiple effect evaporators, resulting in the elimination of the direct contact evaporator.

Kraft Flow Diagram No. 10 represents a new mill designed for a production of 500 to 1000 tons per day.

EMISSIONS

Because there are no mills in the United States operating at present without a direct contact evaporator, Flow Diagram No. 10 illustrates an imaginary mill. Therefore, emissions from the recovery system have been assumed.

POUNDS PER AIR DRY TON OF UNBLEACHED PULP

LOCATION	H_2S	RSH RSR RSSR	SO_2	Particulate
Washers and Screens	0.02	0.24	0	0
High Solids Evaporators	0.50	0.31	0	0
Recovery Boiler	0.10	Trace	5.0	210
After Precipitator	0.10	Trace	5.0	2.10
Smelt Dissolving Tank	Trace	Trace	0	0.20

EMISSIONS - Continued

LOCATION	H_2S	RSH RSR RSSR	SO_2	Particulate
Slaker	0	0	0	Unknown
Lime Kiln	Trace	Trace	Trace	25.0
Lime Kiln Scrubber	Trace	Trace	Trace	0.20
Turpentine Condenser	0.01	0.50	0	0

ASSUMPTIONS

The following assumptions have been made in developing the Flow Diagram:

A. PULP MILL

1. Pulp Yield = 47 Percent

2. Cooking liquor charge per air dry ton of brown stock equals 6,189 pounds of which 889 pounds are chemical solids. Sulfidity equals 26 percent.

B. RECOVERY

1. Unit operating at 15 percent excess air at economizer outlet.

2. Particulate matter of 8 grains per SDCF leaving the economizer.

3. No direct contact evaporators.

4. Steam shatter jets utilized on smelt spouts.

5. Design efficiency of 99.5 percent for precipitator with an annual average operating efficiency of 99.0 percent.

C. CAUSTICIZING

1. Lime kiln scrubber efficiency assumed at 99 percent on lime and 80 percent efficiency on soda.

D. POWER PLANT

1. Excess air: 30 percent for bark, 10 percent for oil, and 20 percent for coal.

2. Bark burning equipment is a low set spreader stoker.

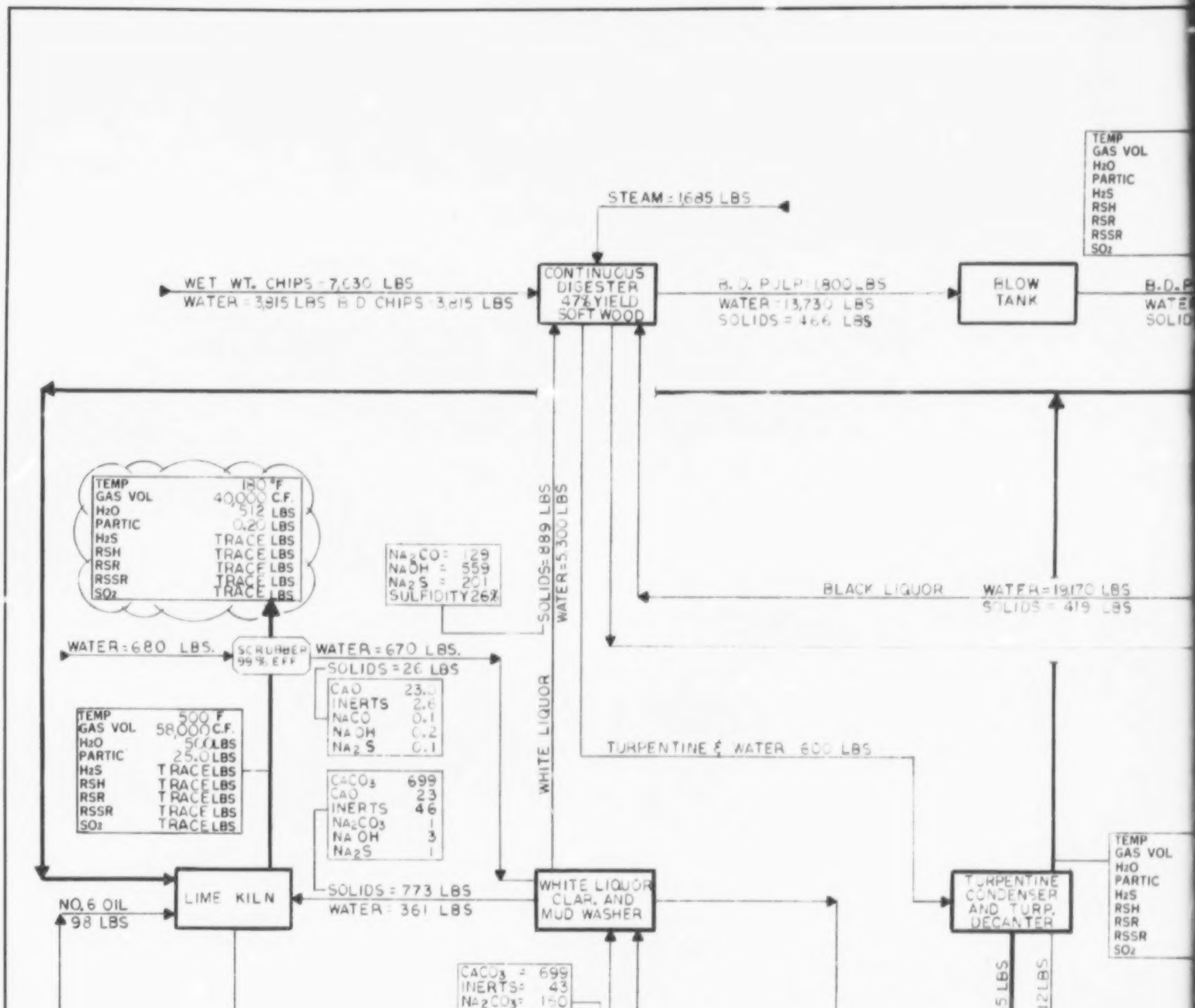
3. Reinjection from dust collector: 50 percent for bark and 0 percent for coal.

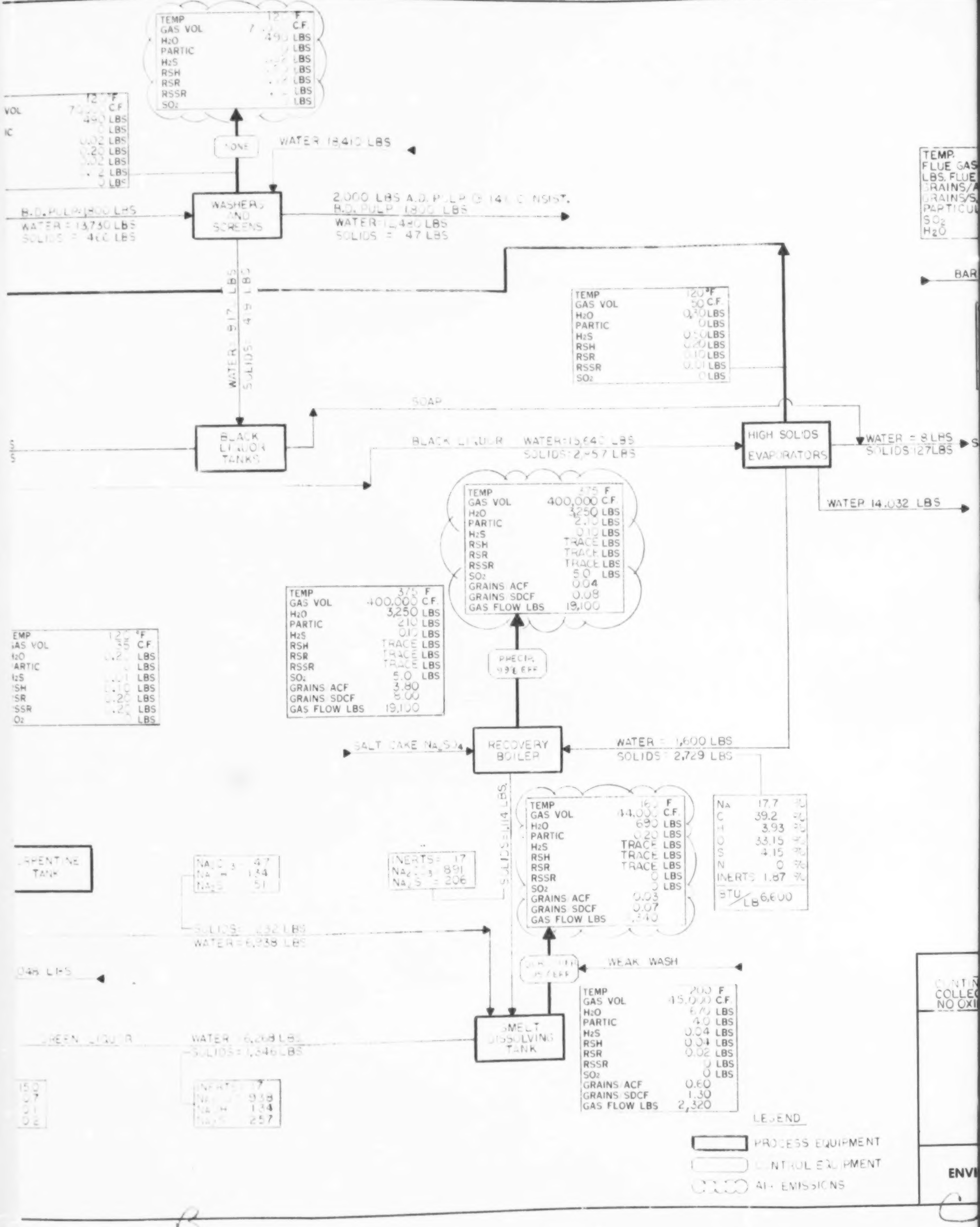
4. Design efficiencies for dust collector: 96 percent on bark and 98 percent on coal with annual operating efficiencies of 94 percent and 96 percent, respectively.

5. Bark is burned 24 hours per day at a controlled rate.

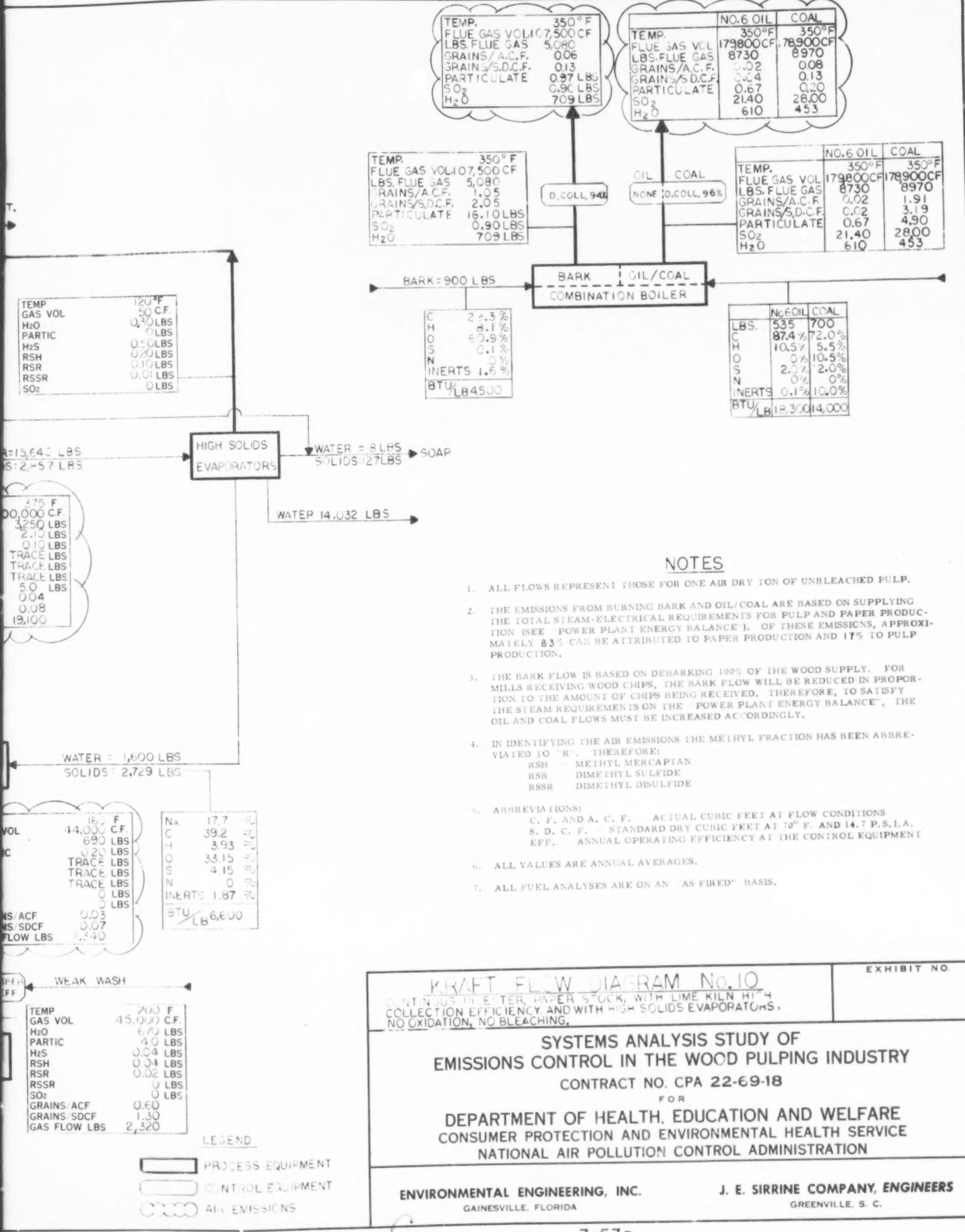
6. Unburned combustible in refuse leaving the dust collector is 40 percent.

7. Coal firing based on pulverized coal.


STEAM = 1685 LBS
WET WT. CHIPS = 7,630 LBS
WATER = 3,815 LBS B.D. CHIPS = 3,815 LBS
CONTINUOUS DIGESTER 47% YIELD SOFT WOOD
B.D. PULP = 1800 LBS
WATER = 13,730 LBS
SOLIDS = 466 LBS
BLOW TANK
B.D.P
WATER
SOLID
TEMP
GAS VOL
H2O
PARTIC
H2S
RSH
RSR
RSSR
SO2
TEMP 180 °F
GAS VOL 40,000 C.F.
H2O 512 LBS
PARTIC 0.20 LBS
H2S TRACE LBS
RSH TRACE LBS
RSR TRACE LBS
RSSR TRACE LBS
SO2 TRACE LBS
NA2CO = 129
NAOH = 559
NA2S = 201
SULFIDITY 26%
SOLIDS = 889 LBS
WATER = 5,300 LBS
WHITE LIQUOR
BLACK LIQUOR
WATER = 19170 LBS
SOLIDS = 419 LBS
WATER = 680 LBS.
SCRUBBER 99% EFF
WATER = 670 LBS.
SOLIDS = 26 LBS
CAO 23.0
INERTS 2.6
NACO 0.1
NA OH 0.2
NA2 S 0.1
TEMP 500 F
GAS VOL 58,000 C.F.
H2O 500 LBS
PARTIC 25.0 LBS
H2S TRACE LBS
RSH TRACE LBS
RSR TRACE LBS
RSSR TRACE LBS
SO2 TRACE LBS
CACO3 699
CAO 23
INERTS 46
NA2CO3 1
NA OH 3
NA2S 1
TURPENTINE & WATER 600 LBS
NO.6 OIL 98 LBS
LIME KILN
SOLIDS = 773 LBS
WATER = 361 LBS
WHITE LIQUOR CLAR. AND MUD WASHER
TURPENTINE CONDENSER AND TURP. DECANTER
TEMP
GAS VOL
H2O
PARTIC
H2S
RSH
RSR
RSSR
SO2
CACO3 = 699
INERTS = 43
NA2CO3 = 160
NAOH = 694
NA2S = 249
SOLIDS = 1845 LBS
WATER = 6575 LBS
WATER = 155 LBS
TURPENTINE = 12 LBS
TURPENTI TANK
TEMP 85 F
GAS VOL 4800 C.F.
H2O LBS
PARTIC UNKNOWN LBS
H2S LBS
RSH LBS
RSR LBS
RSSR LBS
SO2 LBS
SOLIDS = 440 LBS.
TEMP 152 °F
GAS VOL 7, C.F.
H2O 84 LBS
PARTIC UNKNOWN LBS
H2S LBS
RSH LBS
RSR LBS
RSSR LBS
SO2 LBS
SCRUBBER
NA2CO3 = 17
NAOH = 2
NA2S = 4
SOLIDS = 23 LBS
WATER = 5,355 LBS
WATER 76.5 LBS
WATER = 6,048 LBS
CAO 391
INERTS 44
NA2CO3 1
NAOH 3
NA2S 1
SLAKER
WATER = 6,527 LBS
SOLIDS = 1,307 LBS
GREEN LIQUOR CLAR. AND DREGS WASHER
GREE
INERTS = 2
NA2CO3 = 921
NAOH = 131
NA2S = 253
SOLIDS = 15 LBS
CAO = 13.5
INERTS = 1.5
CACO3 = 16.5
INERTS = 3.5
NA2CO3 = 3.0
NAOH 13.0
NA2S = 5.0
SOLIDS = 41 LBS
WATER = 25 LBS
WATER = 134 LBS
SOLIDS = 16 LBS
INERTS = 15.0
NA2CO3 = 0.7
NAOH = 0.1
NA2S = 0.2


TEMP 12_ °F
GAS VOL 7_ C.F.
H2O 490 LBS
PARTIC _ LBS
H2S _ LBS
RSH _ LBS
RSR _ LBS
RSSR _ LBS
SO2 _ LBS
12_ °F
VOL 70__ C.F.
490 LBS
IC 0 LBS
0.02 LBS
0.20 LBS
0.02 LBS
_.2 LBS
0 LBS
NONE
WATER 18,410 LBS
WASHERS AND SCREENS
B.D. PULP 1800 LBS
WATER = 13,730 LBS
SOLIDS = 460 LBS
2,000 LBS A.D. PULP @ 14% CONSIST.
B.D. PULP 1800 LBS
WATER 12,480 LBS
SOLIDS = 47 LBS
WATER 9,17_ LBS
SOLIDS 4_9 LBS
TEMP 120 °F
GAS VOL 50 C.F.
H2O 0.30 LBS
PARTIC 0 LBS
H2S 0.50 LBS
RSH 0.20 LBS
RSR 0.10 LBS
RSSR 0.01 LBS
SO2 0 LBS
SOAP
BLACK LIQUOR TANKS
BLACK LIQUOR WATER 15,640 LBS
SOLIDS 2,957 LBS
HIGH SOLIDS EVAPORATORS
WATER = 8 LBS
SOLIDS 127 LBS
WATER 14,032 LBS
TEMP.
FLUE GAS
LBS. FLUE
GRAINS/A
GRAINS/S
PARTICUL
SO2
H2O
BAR
TEMP 375 °F
GAS VOL 400,000 C.F.
H2O 3,250 LBS
PARTIC 2.10 LBS
H2S 0.10 LBS
RSH TRACE LBS
RSR TRACE LBS
RSSR TRACE LBS
SO2 5.0 LBS
GRAINS/ACF 0.04
GRAINS/SDCF 0.08
GAS FLOW LBS 19,100
TEMP 375 °F
GAS VOL 400,000 C.F.
H2O 3,250 LBS
PARTIC 210 LBS
H2S 0.10 LBS
RSH TRACE LBS
RSR TRACE LBS
RSSR TRACE LBS
SO2 5.0 LBS
GRAINS/ACF 3.80
GRAINS/SDCF 8.00
GAS FLOW LBS 19,100
PRECIP. 99% EFF
TEMP 12_ °F
GAS VOL 35 C.F.
H2O 0.2_ LBS
PARTIC _ LBS
H2S 0.01 LBS
RSH 0.10 LBS
RSR 0.20 LBS
RSSR _.2_ LBS
SO2 _ LBS
SALT CAKE Na2SO4
RECOVERY BOILER
WATER = 1,600 LBS
SOLIDS = 2,729 LBS
SOLIDS = 1,114 LBS
TEMP 160 °F
GAS VOL 44,000 C.F.
H2O 690 LBS
PARTIC 0.20 LBS
H2S TRACE LBS
RSH TRACE LBS
RSR TRACE LBS
RSSR 0 LBS
SO2 0 LBS
GRAINS/ACF 0.03
GRAINS/SDCF 0.07
GAS FLOW LBS 1,340
Na 17.7 %
C 39.2 %
H 3.93 %
O 33.15 %
S 4.15 %
N 0 %
INERTS 1.87 %
BTU/LB 6,600
TURPENTINE TANK
Na2CO3 = 47
NaOH = 134
Na2S = 51
INERTS = 17
Na2CO3 = 891
Na2S = 206
SOLIDS = 232 LBS
WATER = 6,938 LBS
_048 LBS
WEAK WASH
SCRUBBER 85% EFF
TEMP 200 °F
GAS VOL 45,000 C.F.
H2O 670 LBS
PARTIC 4.0 LBS
H2S 0.04 LBS
RSH 0.04 LBS
RSR 0.02 LBS
RSSR 0 LBS
SO2 0 LBS
GRAINS/ACF 0.60
GRAINS/SDCF 1.30
GAS FLOW LBS 2,320
SMELT DISSOLVING TANK
GREEN LIQUOR
WATER = 6,268 LBS
SOLIDS = 1,346 LBS
INERTS = 17
Na2CO3 = 938
NaOH = 134
Na2S = 257
150
07
01
02
CONTIN
COLLEC
NO OXI
ENVI
LEGEND
PROCESS EQUIPMENT
CONTROL EQUIPMENT
AIR EMISSIONS

B


TEMP. 350° F
FLUE GAS VOL.107,500 CF
LBS. FLUE GAS 5,080
GRAINS/A.C.F. 0.06
GRAINS/S.D.C.F. 0.13
PARTICULATE 0.97 LBS
SO_2 0.90 LBS
H_2O 709 LBS
NO.6 OIL | COAL
TEMP. 350°F | 350°F
FLUE GAS VOL 179800CF | 178900CF
LBS. FLUE GAS 8730 | 8970
GRAINS/A.C.F. 0.02 | 0.08
GRAINS/S.D.C.F. 0.04 | 0.13
PARTICULATE 0.67 | 0.20
SO_2 21.40 | 28.00
H_2O 610 | 453
TEMP. 350° F
FLUE GAS VOL.107,500 CF
LBS. FLUE GAS 5,080
GRAINS/A.C.F. 1.05
GRAINS/S.D.C.F. 2.05
PARTICULATE 16.10 LBS
SO_2 0.90 LBS
H_2O 709 LBS
D. COLL. 94%
OIL
COAL
NONE
D. COLL. 96%
NO.6 OIL | COAL
TEMP. 350°F | 350°F
FLUE GAS VOL 179800CF | 178900CF
LBS. FLUE GAS 8730 | 8970
GRAINS/A.C.F. 0.02 | 1.91
GRAINS/S.D.C.F. 0.02 | 3.19
PARTICULATE 0.67 | 4.90
SO_2 21.40 | 28.00
H_2O 610 | 453
T.
BARK = 900 LBS
BARK
OIL/COAL
COMBINATION BOILER
C 23.3 %
H 8.1 %
O 60.9 %
S 0.1 %
N 0 %
INERTS 1.6 %
BTU/LB 4500
No.6 OIL | COAL
LBS. 535 | 700
C 87.4 % | 72.0%
H 10.5% | 5.5%
O 0% | 10.5%
S 2.0% | 2.0%
N 0% | 0%
INERTS 0.1% | 10.0%
BTU/LB 18,300 | 14,000
TEMP 120 °F
GAS VOL 50 C.F.
H_2O 0.30 LBS
PARTIC 0 LBS
H_2S 0.50 LBS
RSH 0.20 LBS
RSR 0.10 LBS
RSSR 0.01 LBS
SO_2 0 LBS
R=15,640 LBS
S: 2,857 LBS
HIGH SOLIDS EVAPORATORS
WATER = 8 LBS
SOLIDS 127 LBS
SOAP
WATER 14,032 LBS
375 F
00,000 C.F.
3,250 LBS
2.10 LBS
0.10 LBS
TRACE LBS
TRACE LBS
TRACE LBS
5.0 LBS
0.04
0.08
19,100
WATER = 1,600 LBS
SOLIDS = 2,729 LBS
160 F
VOL 44,000 C.F.
690 LBS
C 0.20 LBS
TRACE LBS
TRACE LBS
TRACE LBS
0 LBS
0 LBS
NS/ACF 0.03
NS/SDCF 0.07
FLOW LBS 2,340
Na 17.7 %
C 39.2 %
H 3.93 %
O 33.15 %
S 4.15 %
N 0 %
INERTS 1.87 %
BTU/LB 6,600
BBER EFF
WEAK WASH
TEMP 200 F
GAS VOL 45,000 C.F.
H_2O 670 LBS
PARTIC 4.0 LBS
H_2S 0.04 LBS
RSH 0.04 LBS
RSR 0.02 LBS
RSSR 0 LBS
SO_2 0 LBS
GRAINS/ACF 0.60
GRAINS/SDCF 1.30
GAS FLOW LBS 2,320
LEGEND
PROCESS EQUIPMENT
CONTROL EQUIPMENT
AIR EMISSIONS

NOTES

1. ALL FLOWS REPRESENT THOSE FOR ONE AIR DRY TON OF UNBLEACHED PULP.
2. THE EMISSIONS FROM BURNING BARK AND OIL/COAL ARE BASED ON SUPPLYING THE TOTAL STEAM-ELECTRICAL REQUIREMENTS FOR PULP AND PAPER PRODUCTION (SEE "POWER PLANT ENERGY BALANCE"). OF THESE EMISSIONS, APPROXIMATELY 83% CAN BE ATTRIBUTED TO PAPER PRODUCTION AND 17% TO PULP PRODUCTION.
3. THE BARK FLOW IS BASED ON DEBARKING 100% OF THE WOOD SUPPLY. FOR MILLS RECEIVING WOOD CHIPS, THE BARK FLOW WILL BE REDUCED IN PROPORTION TO THE AMOUNT OF CHIPS BEING RECEIVED. THEREFORE, TO SATISFY THE STEAM REQUIREMENTS ON THE "POWER PLANT ENERGY BALANCE", THE OIL AND COAL FLOWS MUST BE INCREASED ACCORDINGLY.
4. IN IDENTIFYING THE AIR EMISSIONS THE METHYL FRACTION HAS BEEN ABBREVIATED TO "R". THEREFORE:
   RSH = METHYL MERCAPTAN
   RSR = DIMETHYL SULFIDE
   RSSR = DIMETHYL DISULFIDE
5. ABBREVIATIONS:
   C. F. AND A. C. F. = ACTUAL CUBIC FEET AT FLOW CONDITIONS
   S. D. C. F. = STANDARD DRY CUBIC FEET AT 70° F. AND 14.7 P.S.I.A.
   EFF. = ANNUAL OPERATING EFFICIENCY AT THE CONTROL EQUIPMENT
6. ALL VALUES ARE ANNUAL AVERAGES.
7. ALL FUEL ANALYSES ARE ON AN "AS FIRED" BASIS.

EXHIBIT NO.

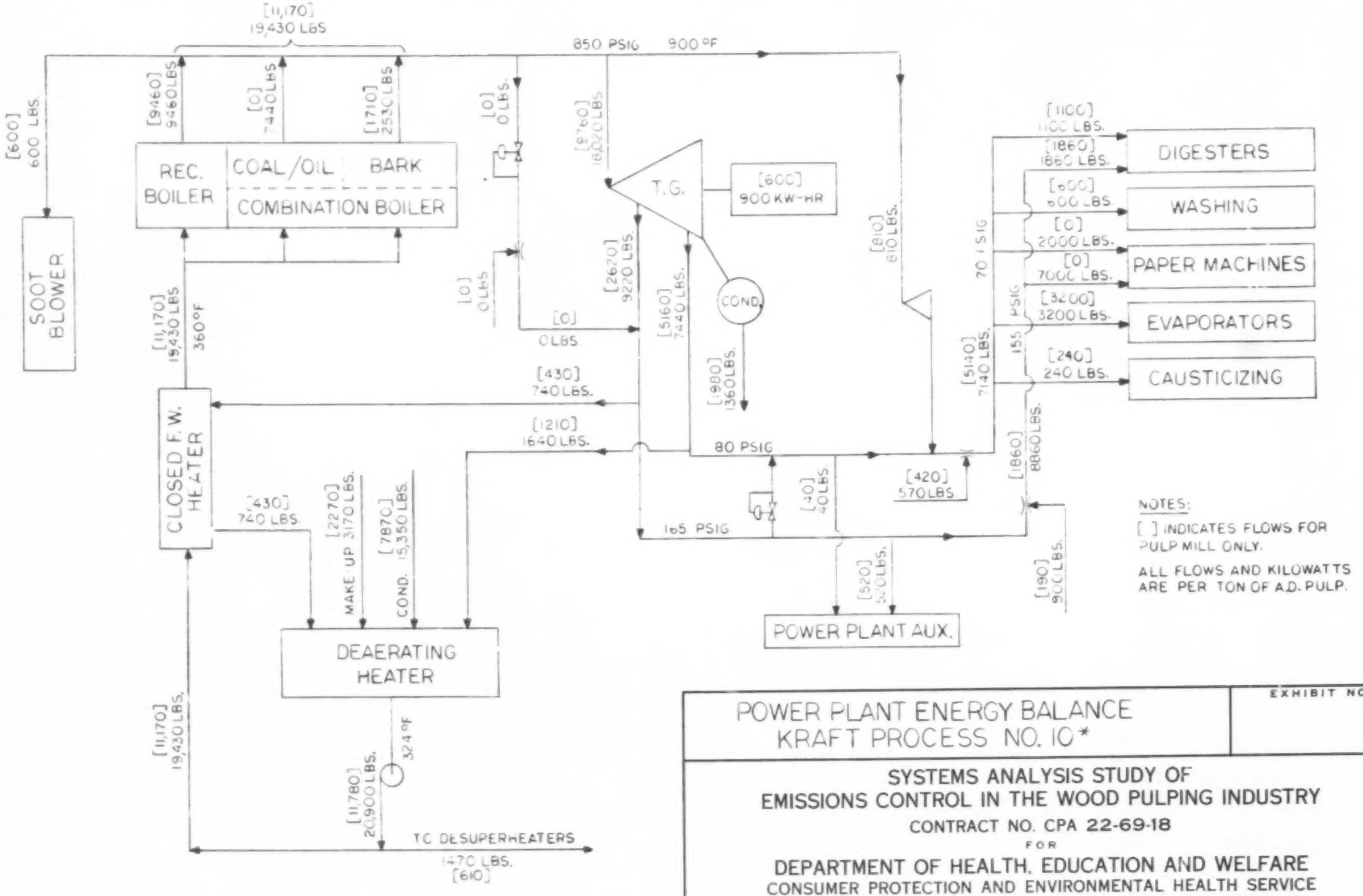
KRAFT FLOW DIAGRAM No.10

CONTINUOUS DIGESTER, PAPER STOCK, WITH LIME KILN HIGH COLLECTION EFFICIENCY, AND WITH HIGH SOLIDS EVAPORATORS, NO OXIDATION, NO BLEACHING.

SYSTEMS ANALYSIS STUDY OF
EMISSIONS CONTROL IN THE WOOD PULPING INDUSTRY
CONTRACT NO. CPA 22-69-18
FOR
DEPARTMENT OF HEALTH, EDUCATION AND WELFARE
CONSUMER PROTECTION AND ENVIRONMENTAL HEALTH SERVICE
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

ENVIRONMENTAL ENGINEERING, INC.
GAINESVILLE, FLORIDA

J. E. SIRRINE COMPANY, *ENGINEERS*
GREENVILLE, S. C.


[11,170]
19,430 LBS
850 PSIG
900 °F
[600]
600 LBS.
SOOT BLOWER
[9460]
9460 LBS
[0]
7440 LBS
[1710]
2530 LBS
REC. BOILER
COAL/OIL
BARK
COMBINATION BOILER
[0]
0 LBS.
[0]
0 LBS
[0]
0 LBS.
[9760]
18,020 LBS
T.G.
[600]
900 KW-HR
[2620]
9220 LBS.
[5160]
7440 LBS.
COND.
[1980]
1360 LBS.
[810]
810 LBS.
[11,170]
19,430 LBS
360°F
CLOSED F.W. HEATER
[430]
740 LBS.
[1210]
1640 LBS.
80 PSIG
[40]
40 LBS.
[420]
570 LBS.
165 PSIG
[520]
520 LBS.
POWER PLANT AUX.
[5140]
7140 LBS.
70 PSIG
155 PSIG
[1860]
8860 LBS.
[190]
900 LBS.
[1100]
1100 LBS.
[1860]
1860 LBS.
DIGESTERS
[600]
600 LBS.
WASHING
[0]
2000 LBS.
[0]
7000 LBS.
PAPER MACHINES
[3200]
3200 LBS.
EVAPORATORS
[240]
240 LBS.
CAUSTICIZING
[430]
740 LBS.
[2270]
MAKE UP 3170 LBS.
[7870]
COND. 15,350 LBS.
DEAERATING HEATER
324 °F
[11,780]
20,900 LBS.
[11,170]
19,430 LBS.
TO DESUPERHEATERS
1470 LBS.
[610]

NOTES:

[ ] INDICATES FLOWS FOR PULP MILL ONLY.

ALL FLOWS AND KILOWATTS ARE PER TON OF A.D. PULP.

* CONTINUOUS DIGESTERS, CONCENTRATED BLACK LIQUOR OXIDATION DIRECT CONTACT EVAPORATOR, NO BLEACHING, HIGH EFFICIENCY LIME KILN COLLECTION

POWER PLANT ENERGY BALANCE
KRAFT PROCESS NO. 10*

EXHIBIT NO.

SYSTEMS ANALYSIS STUDY OF
EMISSIONS CONTROL IN THE WOOD PULPING INDUSTRY
CONTRACT NO. CPA 22-69-18
FOR
DEPARTMENT OF HEALTH, EDUCATION AND WELFARE
CONSUMER PROTECTION AND ENVIRONMENTAL HEALTH SERVICE
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

ENVIRONMENTAL ENGINEERING, INC.
GAINESVILLE, FLORIDA

J. E. SIRRINE COMPANY, *ENGINEERS*
GREENVILLE, S. C.

3.3 NEUTRAL SULFITE SEMICHEMICAL

3.3.1 GENERAL DESCRIPTION OF NSSC

Neutral sulfite semichemical pulping is basically a two-stage process. It involves

1. A mild chemical treatment of the wood chips in the presence of a neutral chemical solution within a digester and followed by

2. A mechanical treatment, called defibering, to disintegrate the wood chips into pulp.

It derives its name "neutral sulfite" from the fact that the solution containing the cooking chemicals, consisting of sodium sulfite and sodium carbonate, is maintained above a pH of 7.0. The name "semichemical" is given because all of the cementing material is not completely removed by the chemical reaction and some mechanical disintegration is required to separate the fibers. Because some of the cementing material remains with the fibers it follows that the "yield" for this process is higher than for a conventional full-chemical pulping process. Semichemical pulping may produce yields of 60 to 80 percent.

The cooking process is carried out in either batch or continuous digesters. Steam maintains the temperature and pressure of the cook within certain limits depending on the end use of the pulp. During this cooking stage odorous gases are created within the digester. At the completion of the cooking cycle, residual pressure within the digester is used to discharge the entire contents of the batch digester into a blow tank. Waste gases, containing the odorous compounds formed in the digester, are usually vented to the atmosphere.

Before the pulp fibers can be used in the production of paper products the spent liquor must be washed from the pulp. This washing is usually performed on multi-stage drum filters. If a kraft system is adjoining, the NSSC spent liquor can be mixed with the spent kraft liquor, up to a limiting percentage, and burned in the recovery furnace. The recovered chemicals are used entirely in the kraft system. Emissions of both sulfur dioxide and hydrogen sulfide may be increased from the kraft recovery furnace when NSSC liquor is added.

Besides the two above spent liquor systems, one discharging the spent liquor to sewer and the other mixing NSSC liquor with kraft liquor, there is a fluidized bed recovery system. This is a patented system in which the NSSC spent liquor is oxidized in a reactor producing a pelleted product, consisting of sodium carbonate and sodium sulfate.

3.3.2 BASIC DESCRIPTION OF NSSC FLOW DIAGRAMS

The basic assumptions which were made in the development of each NSSC flow diagram, the diagram itself, and the energy balance are presented on the following pages.

NSSC DIAGRAM NO. 1

TYPICAL LOCATION

Southern United States

TYPICAL AGE OF EQUIPMENT

Five to ten years.

GENERAL

This flow diagram is based on delivering the NSSC spent liquor to an adjoining kraft recovery system. The emissions shown are those contributed by the NSSC operation only. There are perhaps 10 to 15 mills in the U. S. illustrated by this flow diagram.

EMISSIONS

The addition of NSSC liquor to a kraft recovery system lowers the pH of the kraft liquor. This may result in a greater release of SO_2 and H_2S from the multiple effect and direct contact evaporators.

Valid data are not available to estimate the emissions from the NSSC process steps.

ASSUMPTIONS

The following assumptions have been made in developing the flow diagram:

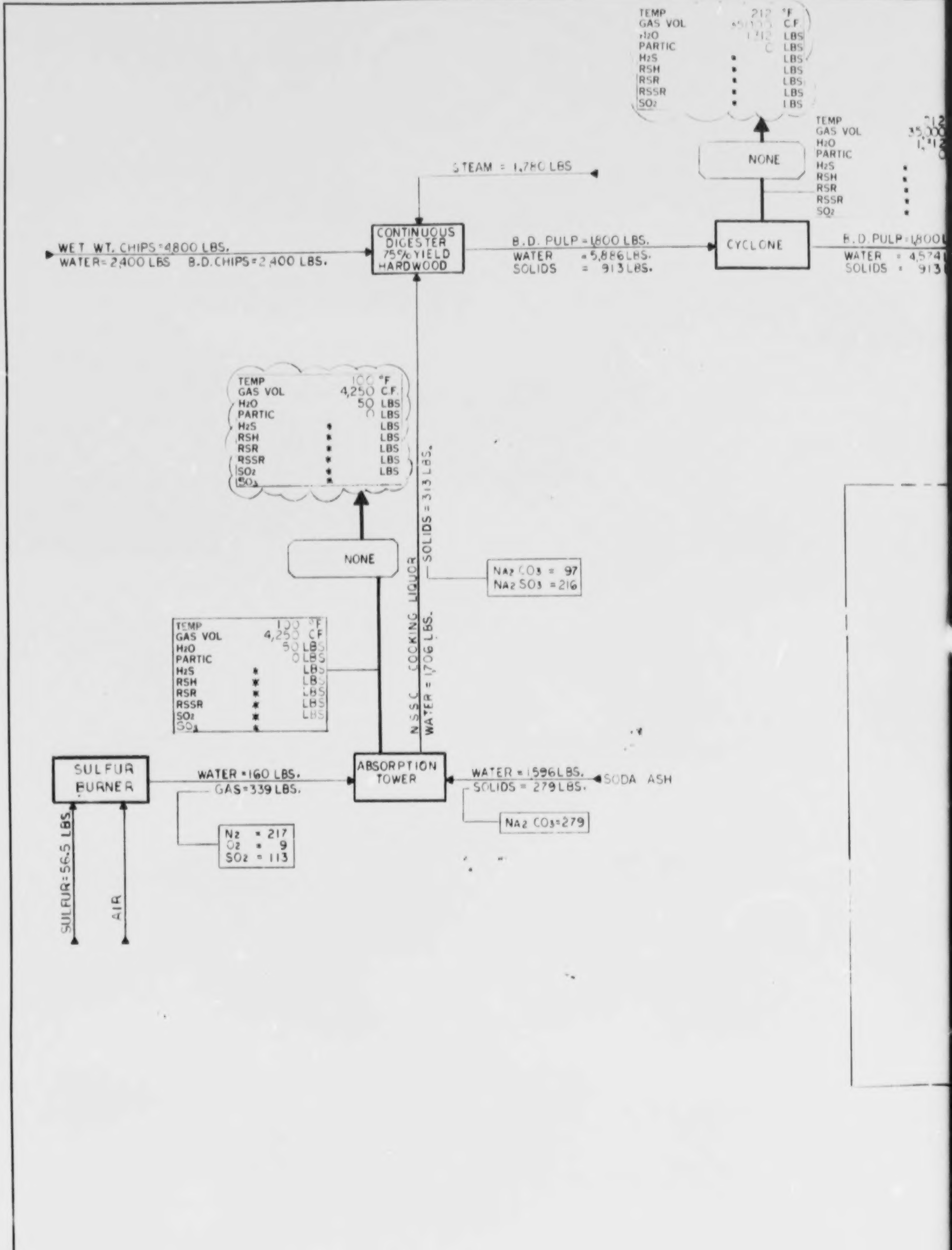
A. RECOVERY

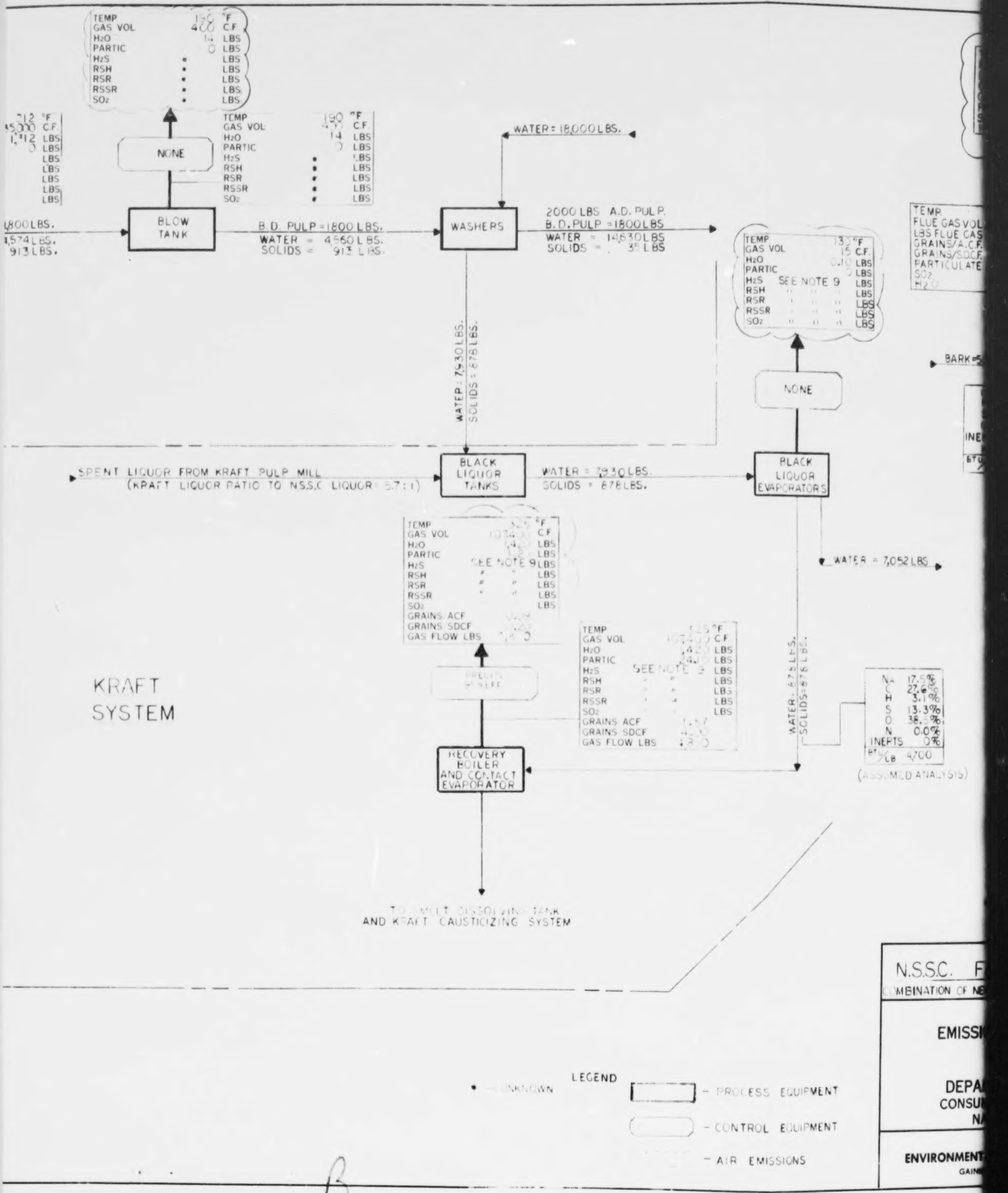
1. Unit operating at 15 percent excess air at economizer outlet.
2. Particulate matter of 8 grains per SDCF leaving the economizer.
3. Efficiency of 50 percent on particulate removal in the direct contact evaporators.
4. Steam shatter jets utilized on smelt spouts.
5. Design efficiency of 98 percent for precipitator with an annual average operating efficiency of 95 percent.

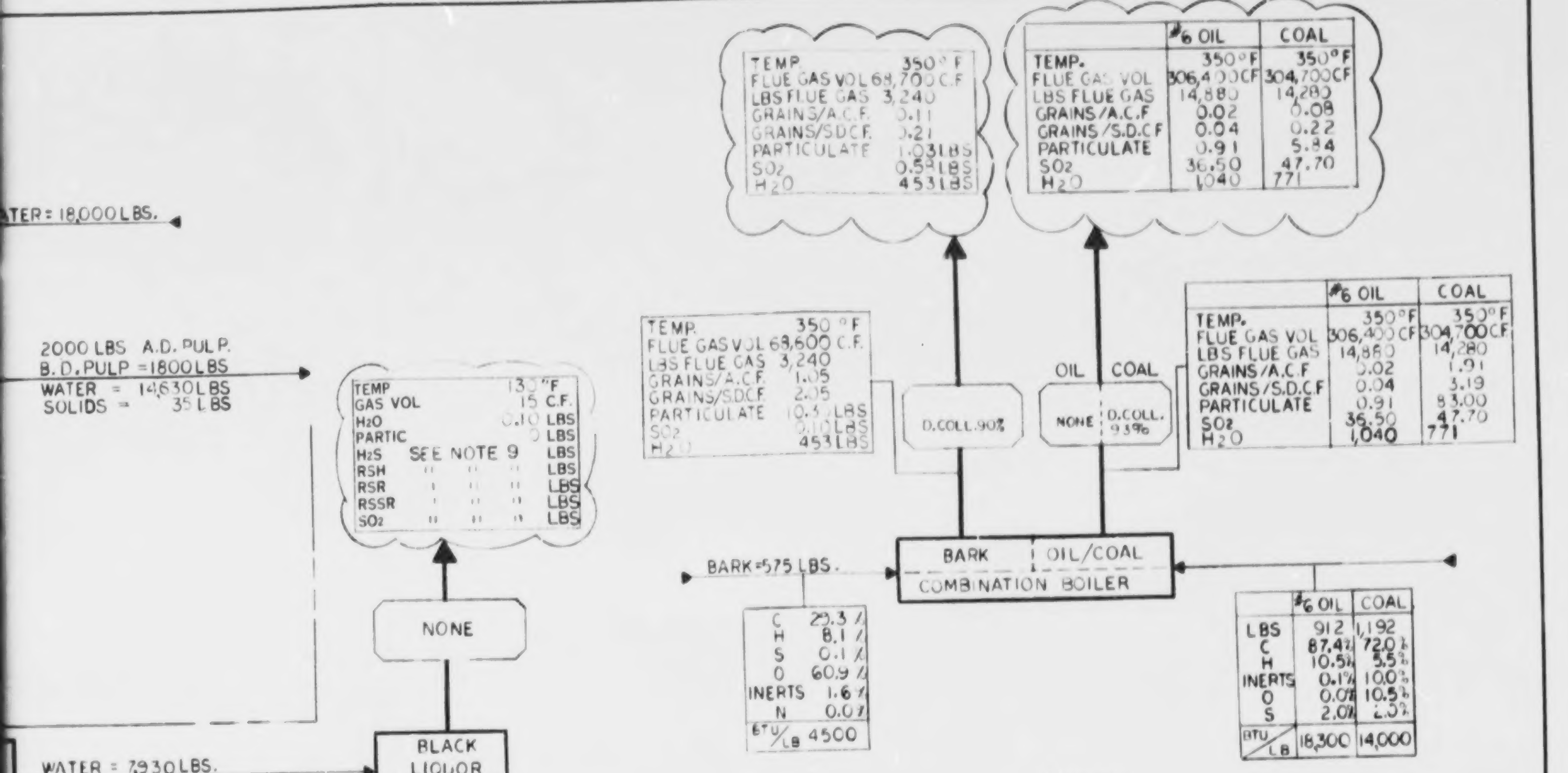
NSSC DIAGRAM NO. 1
(Continued)

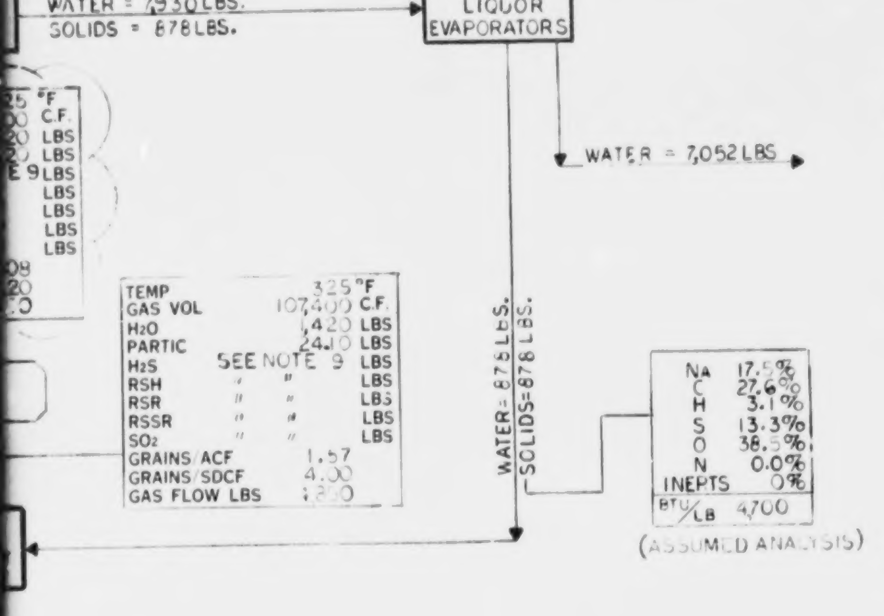
B. POWER PLANT

1. Excess air: 30 percent for bark; 10 percent for oil and 20 percent for coal.
2. Bark burning equipment is a low set spreader stoker.
3. Reinjection from dust collector: 50 percent for bark and 0 percent for coal.
4. Design efficiencies for dust collector: 92 percent on bark and 95 percent on coal with annual operating efficiencies of 90 percent and 93 percent, respectively.
5. Bark is burned 24 hours per day at a controlled rate.
6. Unburned combustible in refuse leaving the dust collector is 40 percent.
7. Coal firing based on pulverized coal.


TEMP 212 °F
GAS VOL 35,000 C.F.
H2O 1,212 LBS
PARTIC 0 LBS
H2S * LBS
RSH * LBS
RSR * LBS
RSSR * LBS
SO2 * LBS
NONE
TEMP 212
GAS VOL 35,000
H2O 1,212
PARTIC 0
H2S *
RSH *
RSR *
RSSR *
SO2 *
STEAM = 1,780 LBS
WET WT. CHIPS = 4800 LBS.
WATER = 2,400 LBS B.D. CHIPS = 2,400 LBS.
CONTINUOUS DIGESTER 75% YIELD HARDWOOD
B.D. PULP = 1800 LBS.
WATER = 5,886 LBS.
SOLIDS = 913 LBS.
CYCLONE
B.D. PULP = 1800
WATER = 4,574
SOLIDS = 913
TEMP 100 °F
GAS VOL 4,250 C.F.
H2O 50 LBS
PARTIC 0 LBS
H2S * LBS
RSH * LBS
RSR * LBS
RSSR * LBS
SO2 * LBS
SO3 *
NONE
N.S.S.C. COOKING LIQUOR
WATER = 1706 LBS.
SOLIDS = 313 LBS.
Na2 CO3 = 97
Na2 SO3 = 216
TEMP 100 °F
GAS VOL 4,250 CF
H2O 50 LBS
PARTIC 0 LBS
H2S * LBS
RSH * LBS
RSR * LBS
RSSR * LBS
SO2 * LBS
SO3 *
SULFUR BURNER
WATER = 160 LBS.
GAS = 339 LBS.
N2 = 217
O2 = 9
SO2 = 113
ABSORPTION TOWER
WATER = 1596 LBS.
SOLIDS = 279 LBS.
SODA ASH
Na2 CO3 = 279
SULFUR = 56.5 LBS.
AIR
A


TEMP 190 °F
GAS VOL 400 C.F.
H2O 14 LBS
PARTIC 0 LBS
H2S • LBS
RSH • LBS
RSR • LBS
RSSR • LBS
SO2 • LBS
212 °F
45,000 C.F.
1,212 LBS
0 LBS
LBS
LBS
LBS
LBS
LBS
NONE
TEMP 190 °F
GAS VOL 400 C.F.
H2O 14 LBS
PARTIC 0 LBS
H2S • LBS
RSH • LBS
RSR • LBS
RSSR • LBS
SO2 • LBS
WATER = 18,000 LBS.
1800 LBS.
4,574 LBS.
913 LBS.
BLOW TANK
B.D. PULP = 1800 LBS.
WATER = 4,560 LBS.
SOLIDS = 913 LBS.
WASHERS
2000 LBS A.D. PULP.
B.D. PULP = 1800 LBS
WATER = 14,630 LBS
SOLIDS = 35 LBS
TEMP 130 °F
GAS VOL 15 C.F.
H2O 0.10 LBS
PARTIC 0 LBS
H2S SEE NOTE 9 LBS
RSH " " " LBS
RSR " " " LBS
RSSR " " " LBS
SO2 " " " LBS
TEMP
FLUE GAS VOL
LBS FLUE GAS
GRAINS/A.C.F.
GRAINS/S.D.C.F.
PARTICULATE
SO2
H2O
WATER: 7,930 LBS.
SOLIDS: 878 LBS.
NONE
BARK=
INE
BTU
SPENT LIQUOR FROM KRAFT PULP MILL
(KRAFT LIQUOR RATIO TO N.S.S.C. LIQUOR= 5.7:1)
BLACK LIQUOR TANKS
WATER = 7,930 LBS.
SOLIDS = 878 LBS.
BLACK LIQUOR EVAPORATORS
WATER = 7,052 LBS
TEMP 325 °F
GAS VOL 107400 C.F.
H2O 1,420 LBS
PARTIC 1.2 LBS
H2S SEE NOTE 9 LBS
RSH " " LBS
RSR " " LBS
RSSR " " LBS
SO2 LBS
GRAINS ACF 0.29
GRAINS SDCF 0.20
GAS FLOW LBS 4,810
TEMP 325 °F
GAS VOL 107,400 C.F.
H2O 1,420 LBS
PARTIC 24.10 LBS
H2S SEE NOTE 9 LBS
RSH " " LBS
RSR " " LBS
RSSR " " LBS
SO2 LBS
GRAINS ACF 1.57
GRAINS SDCF 4.30
GAS FLOW LBS 4,810
PRECIP. 95% EFF.
KRAFT SYSTEM
RECOVERY BOILER AND CONTACT EVAPORATOR
WATER: 878 LBS.
SOLIDS: 878 LBS.
NA 17.5%
C 27.6%
H 3.1%
S 13.3%
O 38.5%
N 0.0%
INERTS 0%
BTU/LB 4700
(ASSUMED ANALYSIS)
TO SMELT DISSOLVING TANK AND KRAFT CAUSTICIZING SYSTEM
N.S.S.C. F
COMBINATION OF NE
EMISSI
DEPA
CONSU
NA
ENVIRONMENT
GAIN
• — UNKNOWN
LEGEND
— PROCESS EQUIPMENT
— CONTROL EQUIPMENT
— AIR EMISSIONS
B


TEMP. 350°F
FLUE GAS VOL 68,700 C.F.
LBS FLUE GAS 3,240
GRAINS/A.C.F. 0.11
GRAINS/SDCF. 0.21
PARTICULATE 1.03 LBS
SO2 0.59 LBS
H2O 453 LBS
#6 OIL | COAL
TEMP. 350°F | 350°F
FLUE GAS VOL 306,400 CF | 304,700 CF
LBS FLUE GAS 14,880 | 14,280
GRAINS/A.C.F 0.02 | 0.08
GRAINS/S.D.C.F 0.04 | 0.22
PARTICULATE 0.91 | 5.84
SO2 36.50 | 47.70
H2O 1040 | 771
ATER = 18,000 LBS.
2000 LBS A.D. PULP.
B.D. PULP = 1800 LBS
WATER = 14,630 LBS
SOLIDS = 35 LBS
TEMP 130°F
GAS VOL 15 C.F.
H2O 0.10 LBS
PARTIC 0 LBS
H2S SEE NOTE 9 LBS
RSH " " " LBS
RSR " " " LBS
RSSR " " " LBS
SO2 " " " LBS
NONE
TEMP. 350°F
FLUE GAS VOL 68,600 C.F.
LBS FLUE GAS 3,240
GRAINS/A.C.F. 1.05
GRAINS/SDCF 2.05
PARTICULATE 10.3 LBS
SO2 0.10 LBS
H2O 453 LBS
D.COLL.90%
OIL
COAL
NONE
D.COLL. 93%
#6 OIL | COAL
TEMP. 350°F | 350°F
FLUE GAS VOL 306,400 CF | 304,700 CF
LBS FLUE GAS 14,880 | 14,280
GRAINS/A.C.F 0.02 | 1.91
GRAINS/S.D.C.F 0.04 | 3.19
PARTICULATE 0.91 | 83.00
SO2 36.50 | 47.70
H2O 1,040 | 771
BARK = 575 LBS.
BARK
OIL/COAL
COMBINATION BOILER
C 29.3%
H 8.1%
S 0.1%
O 60.9%
INERTS 1.6%
N 0.0%
BTU/LB 4500
#6 OIL | COAL
LBS 912 | 1,192
C 87.4% | 72.0%
H 10.5% | 5.5%
INERTS 0.1% | 10.0%
O 0.0% | 10.5%
S 2.0% | 2.0%
BTU/LB 18,300 | 14,000

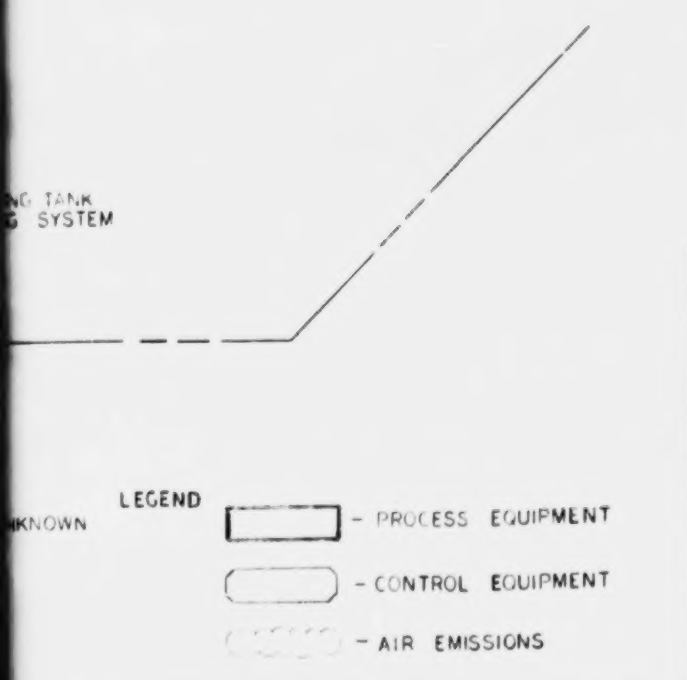

WATER = 7,930 LBS.
SOLIDS = 878 LBS.
BLACK LIQUOR EVAPORATORS
WATER = 7,052 LBS
25 °F
00 C.F.
20 LBS
20 LBS
E 9 LBS
LBS
LBS
LBS
LBS
08
20
.0
TEMP 325°F
GAS VOL 107,400 C.F.
H2O 1,420 LBS
PARTIC 24.10 LBS
H2S SEE NOTE 9 LBS
RSH " " LBS
RSR " " LBS
RSSR " " LBS
SO2 " " LBS
GRAINS/ACF 1.57
GRAINS/SDCF 4.00
GAS FLOW LBS 1,850
WATER = 878 LBS.
SOLIDS = 878 LBS.
NA 17.5%
C 27.6%
H 3.1%
S 13.3%
O 38.5%
N 0.0%
INERTS 0%
BTU/LB 4700
(ASSUMED ANALYSIS)

NOTES

1. ALL FLOWS REPRESENT THOSE FOR ONE AIR DRY TON OF UNBLEACHED PULP.
2. THE EMISSIONS FROM BURNING BARK AND OIL/COAL ARE BASED ON SUPPLYING THE TOTAL STEAM-ELECTRICAL REQUIREMENTS FOR PULP AND PAPER PRODUCTION (SEE "POWER PLANT ENERGY BALANCE"). OF THESE EMISSIONS, APPROXIMATELY 55% CAN BE ATTRIBUTED TO PAPER PRODUCTION AND 45% TO PULP PRODUCTION.
3. THE BARK FLOW IS BASED ON DEBARKING 100% OF THE WOOD SUPPLY. FOR MILLS RECEIVING WOOD CHIPS, THE BARK FLOW WILL BE REDUCED IN PROPORTION TO THE AMOUNT OF CHIPS BEING RECEIVED. THEREFORE, TO SATISFY THE STEAM REQUIREMENTS ON THE "POWER PLANT ENERGY BALANCE", THE OIL AND COAL FLOWS MUST BE INCREASED ACCORDINGLY.
4. IN IDENTIFYING THE AIR EMISSIONS THE METHYL FRACTION HAS BEEN ABBREVIATED TO "R". THEREFORE:

   RSH = METHYL MERCAPTAN
   RSR = DIMETHYL SULFIDE
   RSSR = DIMETHYL DISULFIDE
5. ABBREVIATIONS:

   C. F. AND A. C. F. = ACTUAL CUBIC FEET AT FLOW CONDITIONS
   S. D. C. F. = STANDARD DRY CUBIC FEET AT 70° F. AND 14.7 P.S.I.A.
   EFF. = ANNUAL OPERATING EFFICIENCY AT THE CONTROL EQUIPMENT
6. ALL VALUES ARE ANNUAL AVERAGES.
7. ALL FUEL ANALYSES ARE ON AN "AS FIRED" BASIS.
8. THE EMISSIONS FROM THE RECOVERY BOILER AND BLACK LIQUOR EVAPORATORS ARE BASED ON ONE AIR DRY TON OF N.S.S.C. PULP AND DO NOT INCLUDE ANY EMISSIONS FROM THE KRAFT SYSTEM.
9. RECOVERY BOILER EMISSIONS OF REDUCED SULFUR COMPOUNDS RESULTING FROM THE N.S.S.C. PROCESS ARE UNKNOWN.
10. THE ADDITION OF N.S.S.C. LIQUOR TO A KRAFT RECOVERY SYSTEM LOWERS THE pH OF THE KRAFT LIQUOR. THIS MAY RESULT IN A GREATER RELEASE OF MALODOROUS EMISSIONS FROM THE MULTIPLE EFFECT AND DIRECT CONTACT EVAPORATORS.


NG TANK
G SYSTEM
LEGEND
KNOWN

LEGEND

- ▭ - PROCESS EQUIPMENT
- ▢ - CONTROL EQUIPMENT
- ☁ - AIR EMISSIONS

EXHIBIT NO.

N.S.S.C. FLOW DIAGRAM NO.1

COMBINATION OF NEUTRAL SULFITE SEMI-CHEMICAL AND KRAFT PULPING

SYSTEMS ANALYSIS STUDY OF EMISSIONS CONTROL IN THE WOOD PULPING INDUSTRY

CONTRACT NO. CPA 22-69-18

FOR

DEPARTMENT OF HEALTH, EDUCATION AND WELFARE
CONSUMER PROTECTION AND ENVIRONMENTAL HEALTH SERVICE
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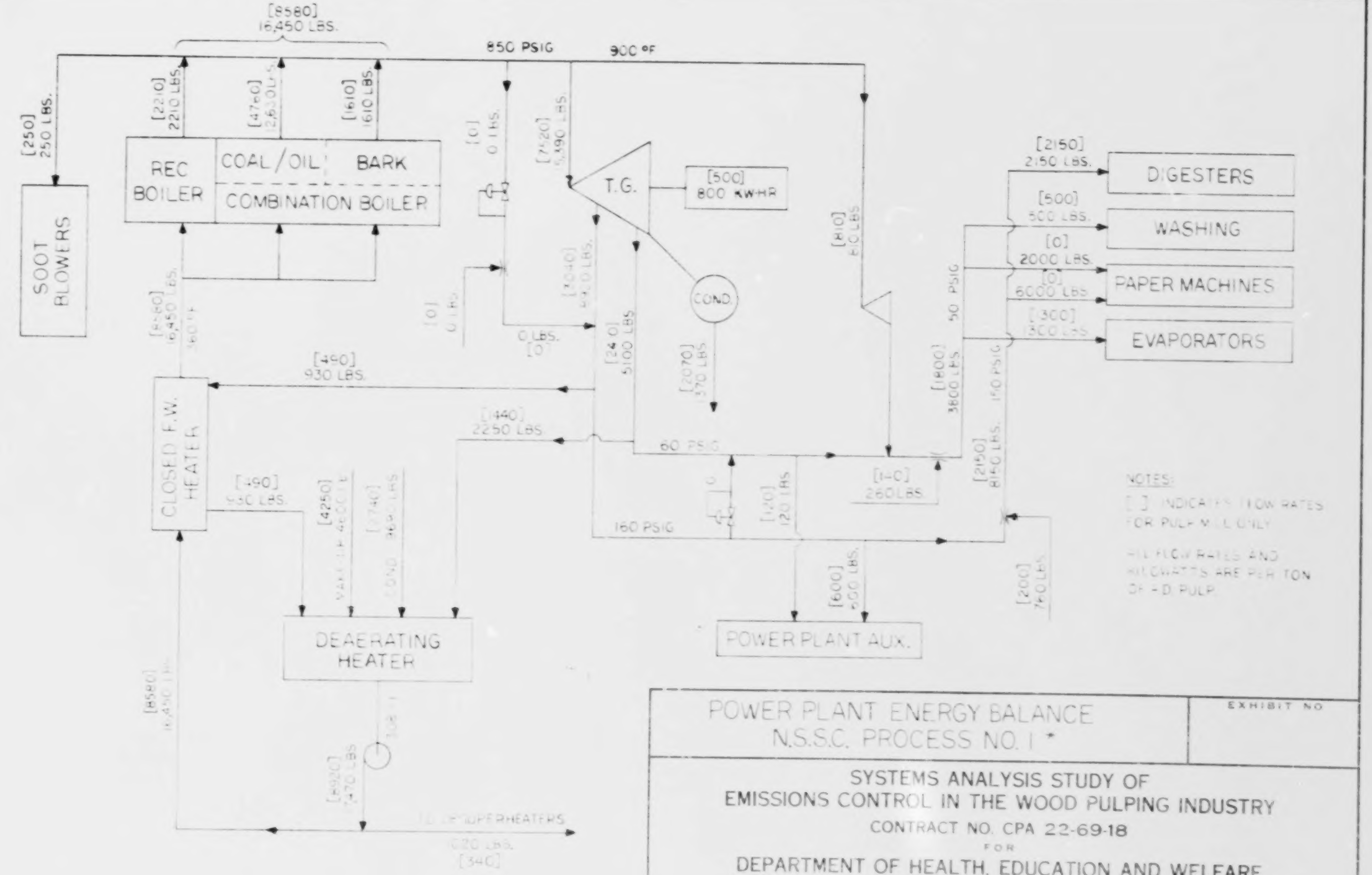
ENVIRONMENTAL ENGINEERING, INC.
GAINESVILLE, FLORIDA

J. E. SIRRINE COMPANY, ENGINEERS
GREENVILLE, S. C.


[8580]
16,450 LBS.
850 PSIG
900 °F
[250]
250 LBS.
SOOT BLOWERS
[2210]
2210 LBS.
[4760]
12,630 LBS.
[1610]
1610 LBS.
REC BOILER
COAL/OIL
BARK
COMBINATION BOILER
[0]
0 LBS.
[7520]
5,390 LBS.
T.G.
[500]
800 KWHR
COND.
[3040]
9400 LBS.
[0]
0 LBS.
0 LBS.
[0]
[2410]
5100 LBS.
[2070]
1370 LBS.
[810]
810 LBS.
[8580]
16,450 LBS.
360 °F
[490]
930 LBS.
[1440]
2250 LBS.
60 PSIG
160 PSIG
0
[120]
120 LBS.
[140]
260 LBS.
[1800]
3800 LBS.
50 PSIG
150 PSIG
[2150]
8150 LBS.
[600]
600 LBS.
[200]
760 LBS.
[2150]
2150 LBS.
DIGESTERS
[500]
500 LBS.
WASHING
[0]
2000 LBS.
[0]
6000 LBS.
PAPER MACHINES
[300]
1300 LBS.
EVAPORATORS
CLOSED F.W. HEATER
[490]
930 LBS.
[4250]
MAKE-UP 4600 LBS.
[2740]
COND. 3690 LBS.
DEAERATING HEATER
POWER PLANT AUX.
[8580]
16,450 LBS.
[8920]
17,470 LBS.
TO DESUPERHEATERS
1020 LBS.
[340]
NOTES:
[ ] INDICATES FLOW RATES FOR PULP MILL ONLY
ALL FLOW RATES AND KILOWATTS ARE PER TON OF A.D. PULP

* COMBINATION OF NEUTRAL SULFITE SEMI-CHEMICAL PULPING AND KRAFT PULPING.

POWER PLANT ENERGY BALANCE
N.S.S.C. PROCESS NO. 1 *

EXHIBIT NO

SYSTEMS ANALYSIS STUDY OF
EMISSIONS CONTROL IN THE WOOD PULPING INDUSTRY
CONTRACT NO. CPA 22-69-18
FOR
DEPARTMENT OF HEALTH, EDUCATION AND WELFARE
CONSUMER PROTECTION AND ENVIRONMENTAL HEALTH SERVICE
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

ENVIRONMENTAL ENGINEERING, INC.
GAINESVILLE, FLORIDA

J. E. SIRRINE COMPANY, *ENGINEERS*
GREENVILLE, S. C.

NSSC DIAGRAM NO. 2

TYPICAL LOCATION

Southern United States.

TYPICAL AGE OF EQUIPMENT

Over fifteen years.

GENERAL

This flow diagram is based on the use of neutral sodium sulfite, without chemical recovery. Cooking is done in batch digesters. There are perhaps 15 to 20 mills in the U. S. illustrated by this flow diagram.

EMISSIONS

Valid data are not available to furnish information on emissions.

ASSUMPTIONS

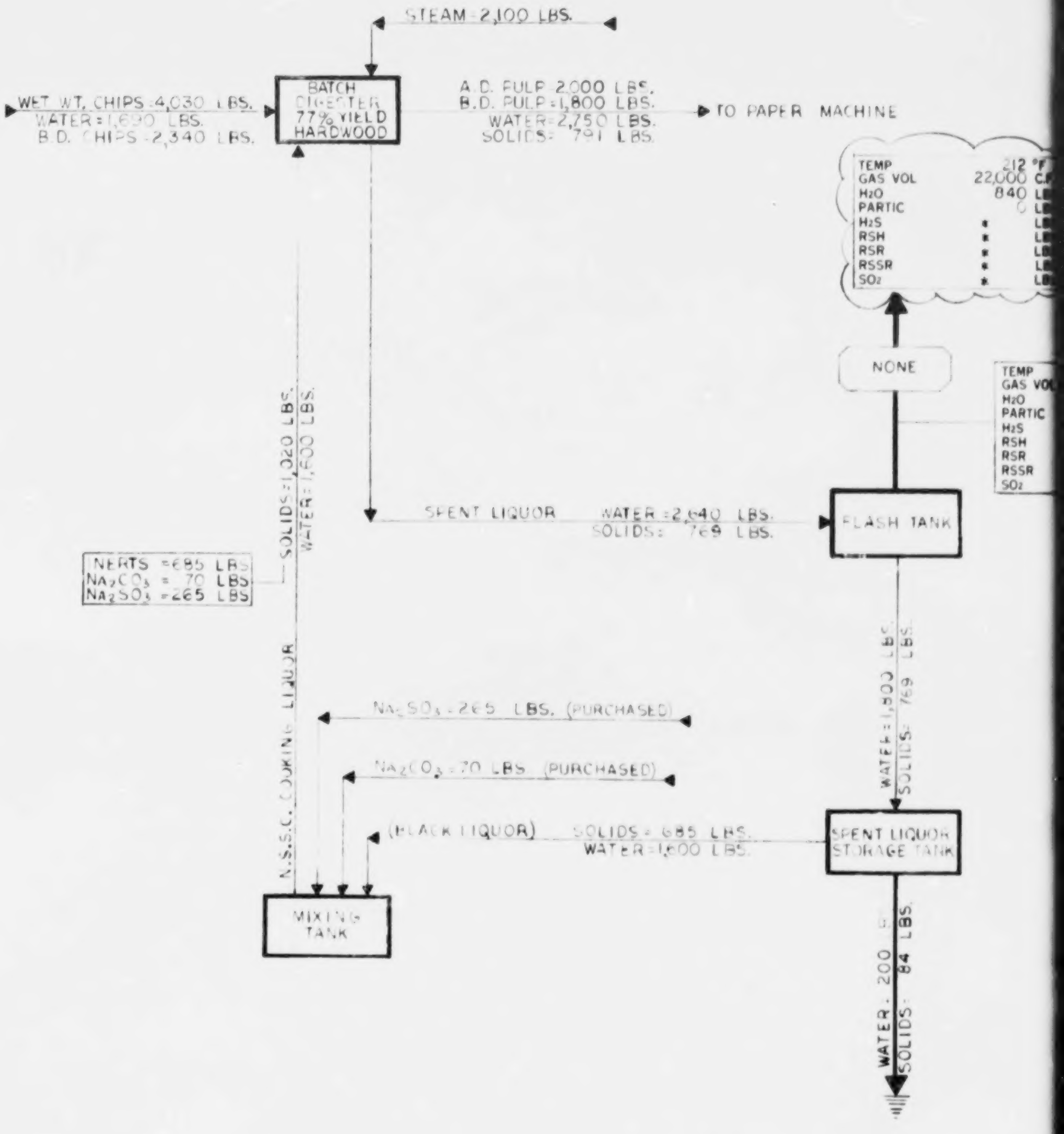
The following assumptions have been made in developing the flow diagram:

A. PULP MILL

1. Pulp Yield = 77 Percent.

B. POWER PLANT

1. Excess air: 30 percent for bark, 10 percent for oil and 20 percent for coal.
2. Bark burning equipment is a low set spreader stoker.
3. Reinjection from dust collector: 50 percent for bark and 0 percent for coal.
4. Design efficiencies for dust collector: 82 percent on bark and 92 percent on coal with annual operating efficiencies of 80 percent and 90 percent, respectively.
5. Bark is burned 24 hours per day at a controlled rate.
6. Unburned combustible in refuse leaving the dust collector is 40 percent.
7. Coal firing based on pulverized coal.


STEAM=2,100 LBS.
WET WT. CHIPS=4,030 LBS.
WATER=1,690 LBS.
B.D. CHIPS=2,340 LBS.
BATCH DIGESTER 77% YIELD HARDWOOD
A.D. PULP 2,000 LBS.
B.D. PULP=1,800 LBS.
WATER=2,750 LBS.
SOLIDS= 791 LBS.
TO PAPER MACHINE
TEMP 212 °F
GAS VOL 22,000 C.
H_2O 840 LB
PARTIC 0 LB
H_2S * LB
RSH * LB
RSR * LB
RSSR * LB
SO_2 * LB
NONE
TEMP
GAS VO
H_2O
PARTIC
H_2S
RSH
RSR
RSSR
SO_2
SOLIDS=1,020 LBS.
WATER=1,600 LBS.
SPENT LIQUOR WATER=2,640 LBS.
SOLIDS= 769 LBS.
FLASH TANK
INERTS =685 LBS
Na_2CO_3 = 70 LBS
Na_2SO_3 =265 LBS
N.S.S.C. COOKING LIQUOR
WATER=1,800 LBS.
SOLIDS= 769 LBS.
Na_2SO_3=265 LBS. (PURCHASED)
Na_2CO_3=70 LBS. (PURCHASED)
(BLACK LIQUOR) SOLIDS=685 LBS.
WATER=1,600 LBS.
SPENT LIQUOR STORAGE TANK
MIXING TANK
WATER= 200 LBS.
SOLIDS= 84 LBS.

A

TE
FU
LB
GR
GR
PA
SO
H_2
TEMP.
FLUE GAS VOL
LBS. FLUE GAS
GRAINS/A.C.F.
GRAINS/S.D.C.F
PARTIC.
SO_2
H_2O
BARK 480
C
H
O
S
N
INE
BTU
22,500
840
°F
C.F
LBS
LBS
*
LBS
*
LBS
*
LBS
*
LBS
*
LBS

LEGEND

* — UNKNOWN

▭ PROCESS EQUIPMENT

( ) - CONTROL EQUIPMENT

☁ AIR EMISSIONS

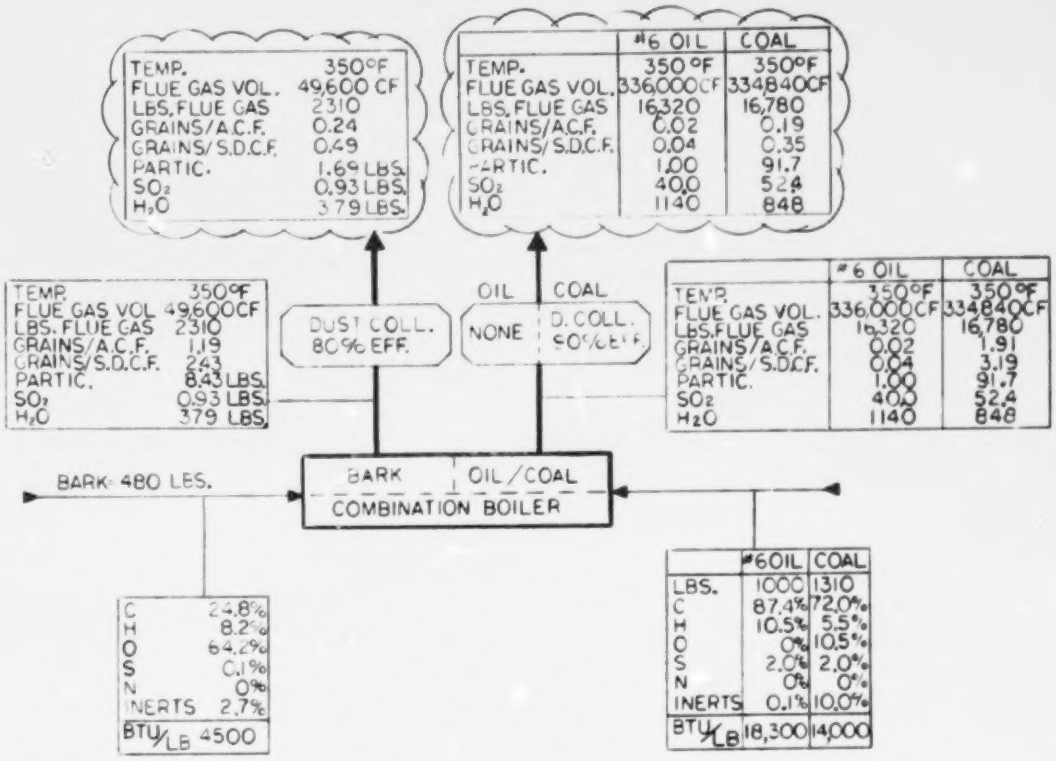
N

BATCH

C

ENVI

B


TEMP. 350°F
FLUE GAS VOL. 49,600 CF
LBS. FLUE GAS 2310
GRAINS/A.C.F. 0.24
GRAINS/S.D.C.F. 0.49
PARTIC. 1.69 LBS.
SO2 0.93 LBS.
H2O 379 LBS.
#6 OIL
COAL
TEMP. 350°F 350°F
FLUE GAS VOL. 336,000CF 334840CF
LBS. FLUE GAS 16,320 16,780
GRAINS/A.C.F. 0.02 0.19
GRAINS/S.D.C.F. 0.04 0.35
PARTIC. 1.00 91.7
SO2 40.0 524
H2O 1140 848
TEMP. 350°F
FLUE GAS VOL 49,600CF
LBS. FLUE GAS 2310
GRAINS/A.C.F. 1.19
GRAINS/S.D.C.F. 2.43
PARTIC. 8.43 LBS.
SO2 0.93 LBS.
H2O 379 LBS.
DUST COLL. 80% EFF.
OIL
COAL
NONE
D. COLL. 90% EFF.
#6 OIL
COAL
TEMP. 350°F 350°F
FLUE GAS VOL. 336,000CF 334840CF
LBS. FLUE GAS 16,320 16,780
GRAINS/A.C.F. 0.02 1.91
GRAINS/S.D.C.F. 0.04 3.19
PARTIC. 1.00 91.7
SO2 40.0 52.4
H2O 1140 848
BARK = 480 LBS.
BARK
OIL/COAL
COMBINATION BOILER
C 24.8%
H 8.2%
O 64.2%
S 0.1%
N 0%
INERTS 2.7%
BTU/LB 4500
#6 OIL
COAL
LBS. 1000 1310
C 87.4% 72.0%
H 10.5% 5.5%
O 0% 10.5%
S 2.0% 2.0%
N 0% 0%
INERTS 0.1% 10.0%
BTU/LB 18,300 14,000

NOTES

1. ALL FLOWS REPRESENT THOSE FOR ONE AIR DRY TON OF UNBLEACHED PULP.
2. THE EMISSIONS FROM BURNING BARK AND OIL/COAL ARE BASED ON SUPPLYING THE TOTAL STEAM-ELECTRICAL REQUIREMENTS FOR PULP AND PAPER PRODUCTION (SEE "POWER PLANT ENERGY BALANCE"). OF THESE EMISSIONS APPROXIMATELY 51% CAN BE ATTRIBUTED TO PAPER PRODUCTION AND 49% TO PULP PRODUCTION.
3. THE BARK FLOW IS BASED ON DEBARKING 100% OF THE WOOD SUPPLY. FOR MILLS RECEIVING WOOD CHIPS, THE BARK FLOW WILL BE REDUCED IN PROPORTION TO THE AMOUNT OF CHIPS BEING RECEIVED. THEREFORE, TO SATISFY THE STEAM REQUIREMENTS ON THE "POWER PLANT ENERGY BALANCE", THE OIL AND COAL FLOWS MUST BE INCREASED ACCORDINGLY.
4. IN IDENTIFYING THE AIR EMISSIONS THE METHYL FRACTION HAS BEEN ABBREVIATED TO "R". THEREFORE:

RSH	METHYL MERCAPTAN
RSR	DIMETHYL SULFIDE
RSSR	DIMETHYL DISULFIDE

5. ABBREVIATIONS:

C. F. AND A. C. F.	ACTUAL CUBIC FEET AT FLOW CONDITIONS
S. D. C. F.	STANDARD DRY CUBIC FEET AT 70° F AND 14.7 P.S.I.A.
EFF.	ANNUAL OPERATING EFFICIENCY AT THE CONTROL EQUIPMENT

6. ALL VALUES ARE ANNUAL AVERAGES.
7. ALL FUEL ANALYSES ARE ON AN "AS FIRED" BASIS.

LEGEND

▭ PROCESS EQUIPMENT

⬭ - CONTROL EQUIPMENT

☁ AIR EMISSIONS

N.S.S.C. FLOW DIAGRAM Nº 2

BATCH DIGESTERS, WITHOUT BLACK LIQUOR RECOVERY

EXHIBIT NO.

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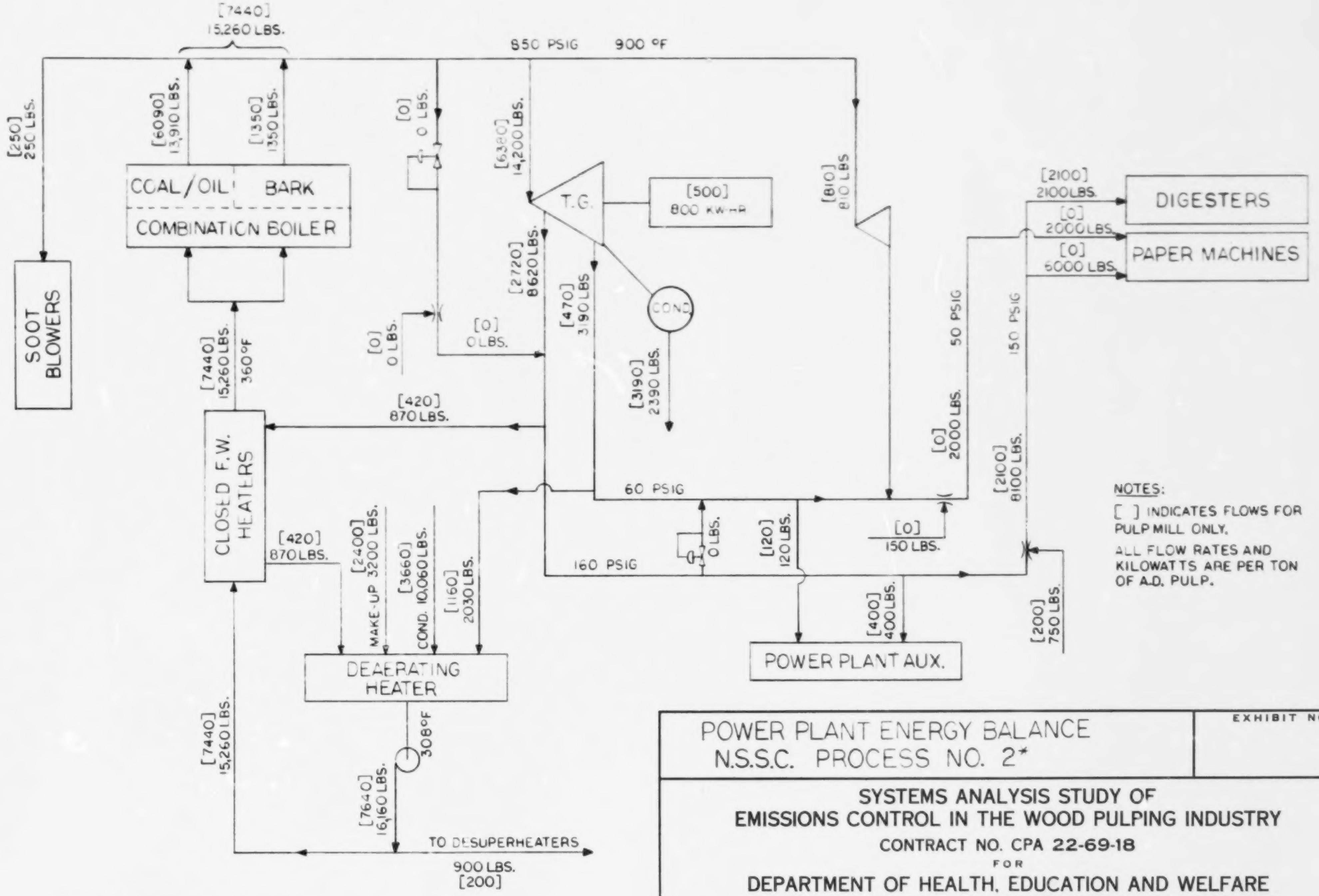
[7440]
15,260 LBS.
850 PSIG 900 °F
[250]
250 LBS.
[6090]
13,910 LBS.
[1350]
1350 LBS.
COAL/OIL
BARK
COMBINATION BOILER
SOOT BLOWERS
[0]
0 LBS.
[6380]
14,200 LBS.
T.G.
[500]
800 KW-HR
[810]
810 LBS.
[2100]
2100 LBS.
DIGESTERS
[0]
2000 LBS.
[0]
6000 LBS.
PAPER MACHINES
[2720]
8620 LBS.
[470]
3190 LBS.
COND.
[3190]
2390 LBS.
[0]
0 LBS.
[0]
0 LBS.
[7440]
15,260 LBS.
360°F
[420]
870 LBS.
50 PSIG
150 PSIG
[0]
2000 LBS.
[2100]
8100 LBS.
CLOSED F.W. HEATERS
60 PSIG
[420]
870 LBS.
[2400]
MAKE-UP 3200 LBS.
[3660]
COND. 10,060 LBS.
[1160]
2030 LBS.
0 LBS.
[120]
120 LBS.
[0]
150 LBS.
160 PSIG
[400]
400 LBS.
[200]
750 LBS.
POWER PLANT AUX.
DEAERATING HEATER
308°F
[7440]
15,260 LBS.
[7640]
16,160 LBS.
TO DESUPERHEATERS
900 LBS.
[200]

NOTES:

[ ] INDICATES FLOWS FOR PULP MILL ONLY.

ALL FLOW RATES AND KILOWATTS ARE PER TON OF A.D. PULP.

* BATCH DIGESTERS WITHOUT BLACK LIQUOR RECOVERY.

POWER PLANT ENERGY BALANCE
N.S.S.C. PROCESS NO. 2*

EXHIBIT NO.

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GAINESVILLE, FLORIDA

J. E. SIRRINE COMPANY, *ENGINEERS*
GREENVILLE, S. C.

NSSC DIAGRAM NO. 3

TYPICAL LOCATION

Midwestern United States.

TYPICAL AGE OF EQUIPMENT

Ten to fifteen years.

GENERAL

This flow diagram is based on the use of neutral sodium sulfite with chemical recovery by the fluidized-bed process. Cooking is done in batch digesters. There are perhaps 2 to 4 mills in the U. S. which are illustrated by this flow diagram.

EMISSIONS

Valid data are not available to present information on emissions except as shown on the flow diagram.

ASSUMPTIONS

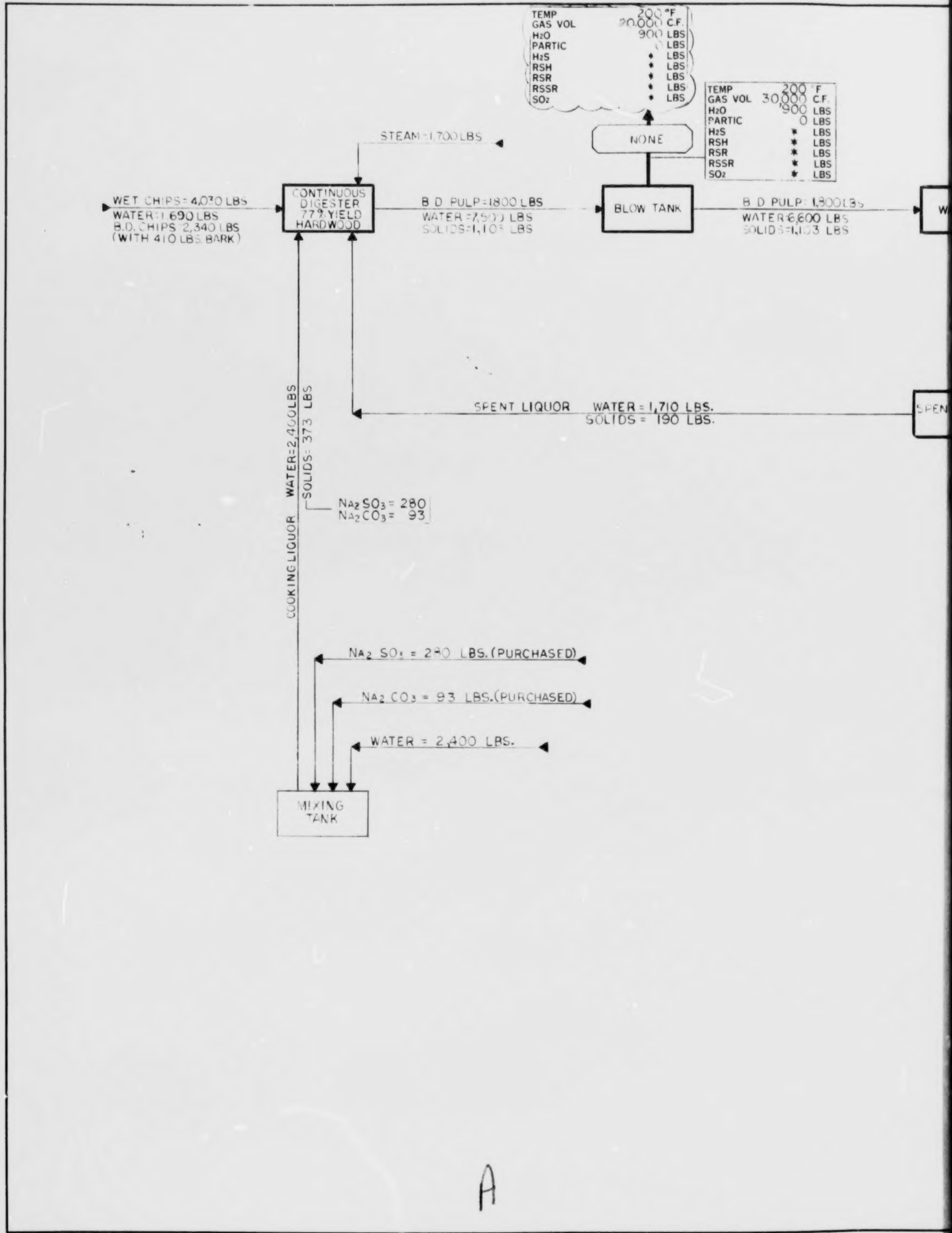
The following assumptions have been made in developing the flow diagram:

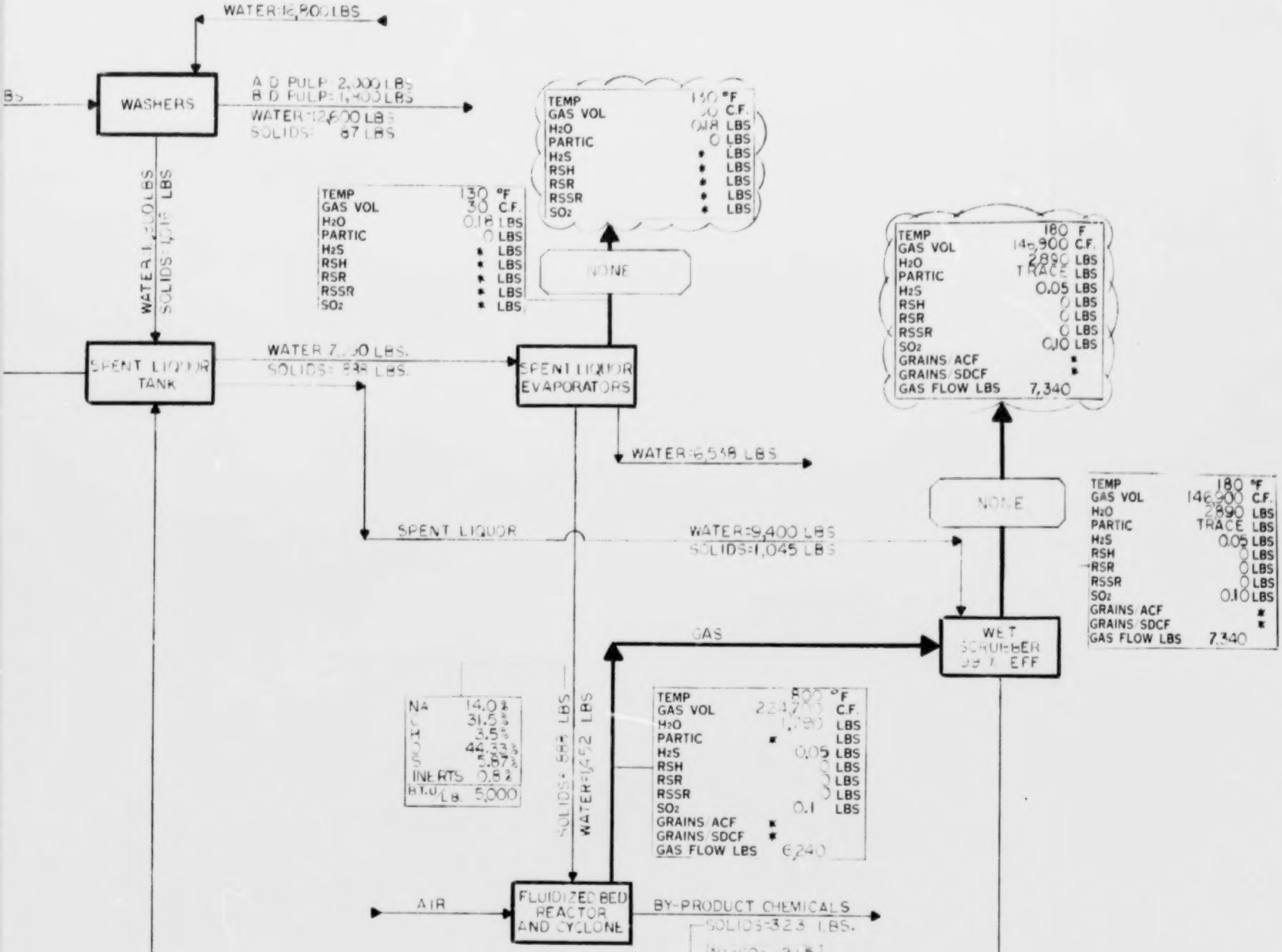
A. PULP MILL

1. Pulp Yield = 77 Percent.

B. POWER PLANT

1. Excess Air: 10 percent for oil and 20 percent for coal.
2. Reinjection from dust collector: 0 percent.
3. Design efficiencies for dust collector: 96 percent on coal with annual operating efficiency of 95 percent.
4. Coal firing based on pulverized coal.
5. Roundwood is not debarked. Bark goes to the digester with the wood.


TEMP 200 °F
GAS VOL 20,000 C.F.
H2O 900 LBS
PARTIC 0 LBS
H2S * LBS
RSH * LBS
RSR * LBS
RSSR * LBS
SO2 * LBS
TEMP 200 °F
GAS VOL 30,000 C.F.
H2O 900 LBS
PARTIC 0 LBS
H2S * LBS
RSH * LBS
RSR * LBS
RSSR * LBS
SO2 * LBS
NONE
STEAM=1,700 LBS
WET CHIPS=4,030 LBS
WATER=1,690 LBS
B.D. CHIPS=2,340 LBS
(WITH 410 LBS. BARK)
CONTINUOUS DIGESTER 77% YIELD HARDWOOD
B D PULP=1800 LBS
WATER=7,500 LBS
SOLIDS=1,103 LBS
BLOW TANK
B D PULP=1,800 LBS
WATER=6,600 LBS
SOLIDS=1,103 LBS
W
SPENT LIQUOR WATER=1,710 LBS.
SOLIDS= 190 LBS.
SPEN
COOKING LIQUOR WATER=2,400 LBS
SOLIDS= 373 LBS
Na2SO3= 280
Na2CO3= 93
NA2 SO3 = 280 LBS. (PURCHASED)
NA2 CO3 = 93 LBS. (PURCHASED)
WATER = 2,400 LBS.
MIXING TANK
A



WATER:12,800 LBS
BS
WASHERS
A D PULP: 2,000 LBS
B D PULP: 1,800 LBS
WATER: 12,600 LBS
SOLIDS: 87 LBS
WATER: 1,300 LBS
SOLIDS: 1,016 LBS
TEMP 130 °F
GAS VOL 30 C.F.
H2O 0.18 LBS
PARTIC 0 LBS
H2S * LBS
RSH * LBS
RSR * LBS
RSSR * LBS
SO2 * LBS
NONE
TEMP 130 °F
GAS VOL 30 C.F.
H2O 0.18 LBS
PARTIC 0 LBS
H2S * LBS
RSH * LBS
RSR * LBS
RSSR * LBS
SO2 * LBS
SPENT LIQUOR TANK
WATER 7,350 LBS.
SOLIDS: 888 LBS.
SPENT LIQUOR EVAPORATORS
WATER: 6,538 LBS
TEMP 180 F
GAS VOL 146,900 C.F.
H2O 2890 LBS
PARTIC TRACE LBS
H2S 0.05 LBS
RSH 0 LBS
RSR 0 LBS
RSSR 0 LBS
SO2 0.10 LBS
GRAINS/ACF *
GRAINS/SDCF *
GAS FLOW LBS 7,340
NONE
TEMP 180 °F
GAS VOL 146,900 C.F.
H2O 2890 LBS
PARTIC TRACE LBS
H2S 0.05 LBS
RSH 0 LBS
RSR 0 LBS
RSSR 0 LBS
SO2 0.10 LBS
GRAINS/ACF *
GRAINS/SDCF *
GAS FLOW LBS 7,340
SPENT LIQUOR
WATER: 9,400 LBS
SOLIDS: 1,045 LBS
GAS
WET SCRUBBER 99 % EFF
NA 14.0%
C 31.5%
H 3.5%
O 44.33%
S 5.87%
INERTS 0.8%
B.T.U./LB. 5,000
SOLIDS: 888 LBS
WATER: 1,452 LBS
TEMP 800 °F
GAS VOL 224,700 C.F.
H2O 1,790 LBS
PARTIC * LBS
H2S 0.05 LBS
RSH 0 LBS
RSR 0 LBS
RSSR 0 LBS
SO2 0.1 LBS
GRAINS/ACF *
GRAINS/SDCF *
GAS FLOW LBS 6,240
AIR
FLUIDIZED BED REACTOR AND CYCLONE
BY-PRODUCT CHEMICALS
SOLIDS: 323 LBS.
NA2SO4 215
NA2CO3 106
INERTS 2
WATER: 8,300 LBS
SOLIDS: 1,107 LBS

N.S.S

FLU

EMISS

DEPA

CONSU

N

ENVIRONMEN

GAIN

C

* — UNKNOWN

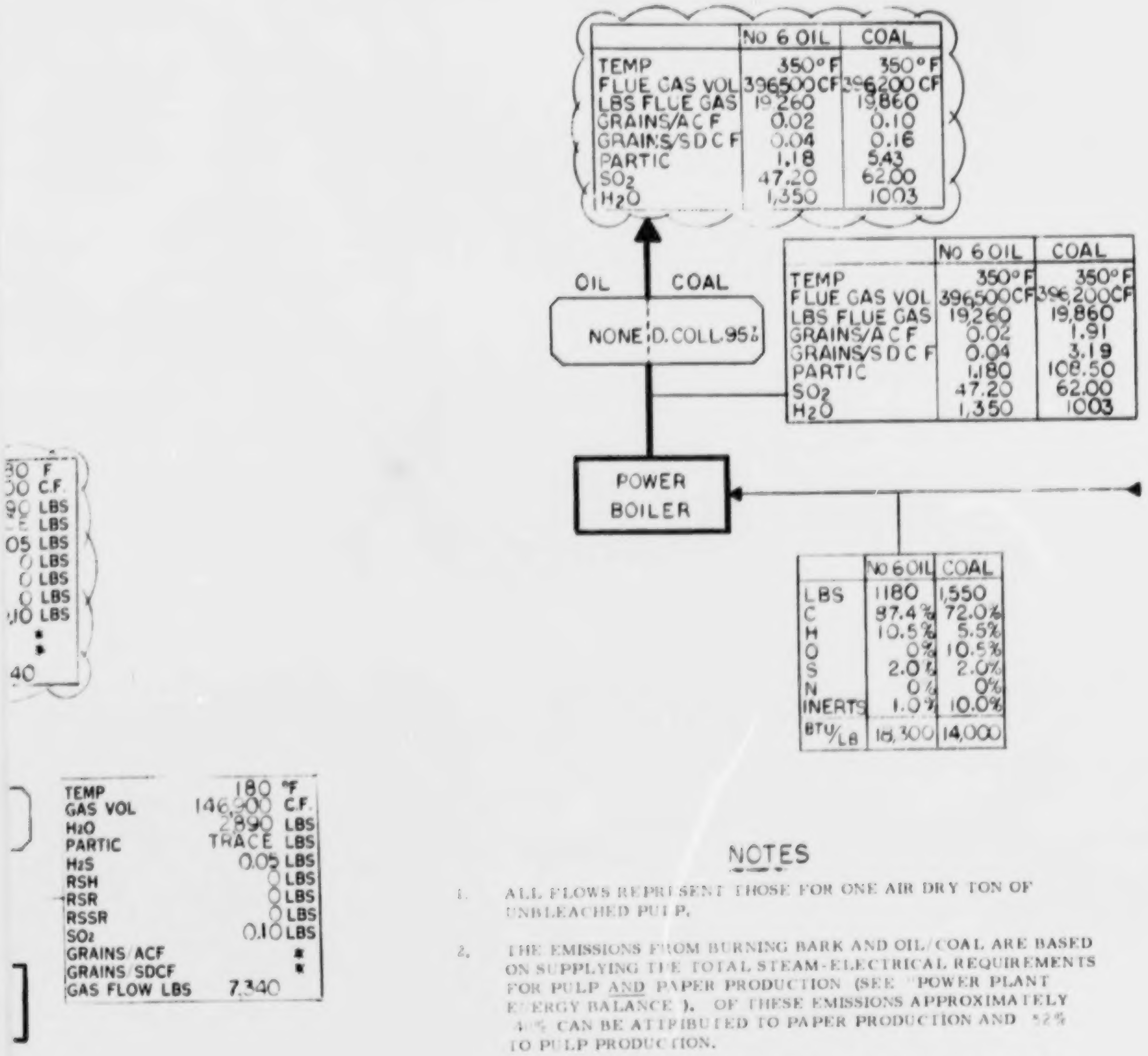
LEGEND

▭ PROCESS EQUIPMENT

⬭ CONTROL EQUIPMENT

☁ AIR EMISSIONS

B



	No 6 OIL	COAL
TEMP	350° F	350° F
FLUE GAS VOL	396500 CF	396200 CF
LBS FLUE GAS	19,260	19,860
GRAINS/A C F	0.02	0.10
GRAINS/S D C F	0.04	0.16
PARTIC	1.18	5.43
SO_2	47.20	62.00
H_2O	1,350	1003

	No 6 OIL	COAL
TEMP	350° F	350° F
FLUE GAS VOL	396,500 CF	396,200 CF
LBS FLUE GAS	19,260	19,860
GRAINS/A C F	0.02	1.91
GRAINS/S D C F	0.04	3.19
PARTIC	1.180	108.50
SO_2	47.20	62.00
H_2O	1,350	1003

	No 6 OIL	COAL
LBS	1180	1,550
C	87.4%	72.0%
H	10.5%	5.5%
O	0%	10.5%
S	2.0%	2.0%
N	0%	0%
INERTS	1.0%	10.0%
BTU/LB	18,300	14,000

30	F
00	C.F.
90	LBS
	LBS
05	LBS
0	LBS
0	LBS
0	LBS
10	LBS
*	
*	
40	

TEMP	180	°F
GAS VOL	146,900	C.F.
H_2O	2890	LBS
PARTIC	TRACE	LBS
H_2S	0.05	LBS
RSH	0	LBS
RSR	0	LBS
RSSR	0	LBS
SO_2	0.10	LBS
GRAINS/ACF		*
GRAINS/SDCF		*
GAS FLOW LBS	7,340	

NOTES

1. ALL FLOWS REPRESENT THOSE FOR ONE AIR DRY TON OF UNBLEACHED PULP.
2. THE EMISSIONS FROM BURNING BARK AND OIL/COAL ARE BASED ON SUPPLYING THE TOTAL STEAM-ELECTRICAL REQUIREMENTS FOR PULP AND PAPER PRODUCTION (SEE "POWER PLANT ENERGY BALANCE"). OF THESE EMISSIONS APPROXIMATELY 48% CAN BE ATTRIBUTED TO PAPER PRODUCTION AND 52% TO PULP PRODUCTION.
3. THE BARK FLOW IS BASED ON DEBARKING 100% OF THE WOOD SUPPLY. FOR MILLS RECEIVING WOOD CHIPS, THE BARK FLOW WILL BE REDUCED IN PROPORTION TO THE AMOUNT OF CHIPS BEING RECEIVED. THEREFORE, TO SATISFY THE STEAM REQUIREMENTS ON THE "POWER PLANT ENERGY BALANCE", THE OIL AND COAL FLOWS MUST BE INCREASED ACCORDINGLY.
4. IN IDENTIFYING THE AIR EMISSIONS THE METHYL FRACTION HAS BEEN ABBREVIATED TO "R". THEREFORE:
   RSH = METHYL MERCAPTAN
   RSR = DIMETHYL SULFIDE
   RSSR = DIMETHYL DISULFIDE
5. ABBREVIATIONS:
   C. F. AND A. C. F. = ACTUAL CUBIC FEET AT FLOW CONDITIONS
   S. D. C. F. = STANDARD DRY CUBIC FEET AT 70° F. AND 14.7 P.S.I.A.
   EFF. = ANNUAL OPERATING EFFICIENCY AT THE CONTROL EQUIPMENT
6. ALL VALUES ARE ANNUAL AVERAGES.
7. ALL FUEL ANALYSES ARE ON AN "AS FIRED" BASIS.

ND
ESS EQUIPMENT
ROL EQUIPMENT
EMISSIONS

C

N.S.S.C. FLOW DIAGRAM No. 3

FLUIDIZED-BED RECOVERY PROCESS

EXHIBIT NO.

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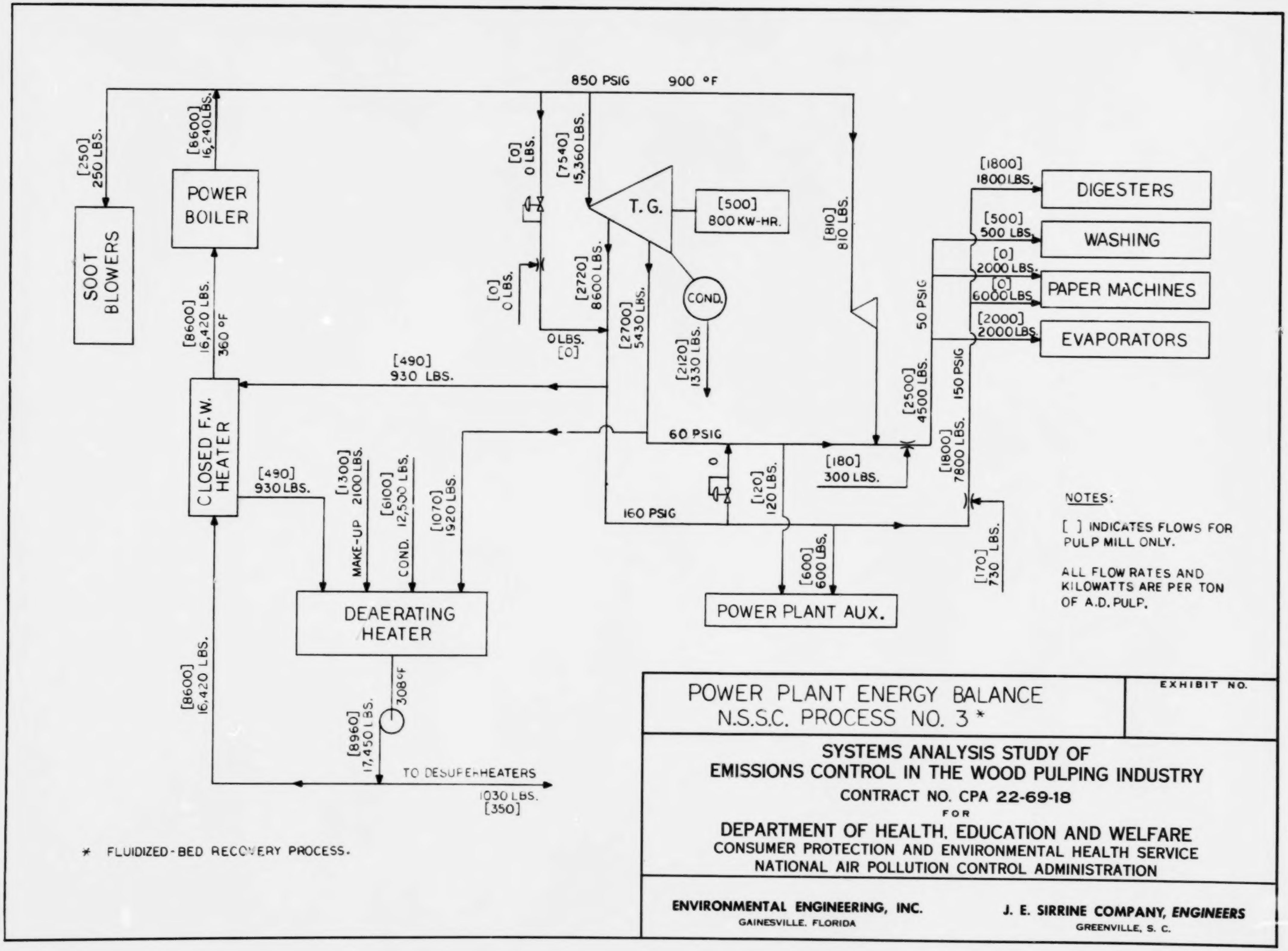
850 PSIG
900 °F
[250]
250 LBS.
SOOT BLOWERS
[8600]
16,240 LBS.
POWER BOILER
[8600]
16,420 LBS.
360 °F
CLOSED F.W. HEATER
[490]
930 LBS.
[490]
930 LBS.
[1300]
2100 LBS.
MAKE-UP
[6100]
12,500 LBS.
COND.
[1070]
1920 LBS.
DEAERATING HEATER
308 °F
[8960]
17,450 LBS.
TO DESUPERHEATERS
1030 LBS.
[350]
[8600]
16,420 LBS.
[0]
0 LBS.
[0]
0 LBS.
0 LBS.
[0]
[7540]
15,360 LBS.
T. G.
[500]
800 KW-HR.
[2720]
8600 LBS.
[2700]
5430 LBS.
COND.
[2120]
1330 LBS.
60 PSIG
160 PSIG
0
[120]
120 LBS.
[600]
600 LBS.
POWER PLANT AUX.
[810]
810 LBS.
[180]
300 LBS.
[2500]
4500 LBS.
50 PSIG
[1800]
7800 LBS.
150 PSIG
[170]
730 LBS.
[1800]
1800 LBS.
DIGESTERS
[500]
500 LBS.
WASHING
[0]
2000 LBS.
[0]
6000 LBS.
PAPER MACHINES
[2000]
2000 LBS.
EVAPORATORS

NOTES:

[ ] INDICATES FLOWS FOR PULP MILL ONLY.

ALL FLOW RATES AND KILOWATTS ARE PER TON OF A.D. PULP.

* FLUIDIZED-BED RECOVERY PROCESS.

POWER PLANT ENERGY BALANCE
N.S.S.C. PROCESS NO. 3 *

EXHIBIT NO.

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GREENVILLE, S. C.

3.4 SULFITE

3.4.1 GENERAL DESCRIPTION OF SULFITE PROCESS

Sulfite pulping is an acid chemical method of dissolving the lignin that bonds the cellulose fibers together. Many of the older mills use a sulfurous acid - calcium bisulfite solution for the cooking acid. Calcium-base spent liquor, because of problems associated with evaporation and chemical recovery, is discarded and may result in water pollution problems. In order to overcome the problem of water pollution several other acid bases have been developed, the most important being sodium, magnesium, and ammonium.

Because sulfite pulp is used in a wide variety of end products, operations will vary considerably between mills. These products can include pulp for making high grade book and bond papers, tissues, for combining with other pulps, and for making dissolving pulp for producing cellophane, rayon, acetate, films, and others.

The pulping operation involves cooking the wood chips in the presence of an acid within a digester. The heat required for cooking is produced by the direct addition of steam to the digester or by the steam heating of the recirculated acid in an external heat exchanger. The cooking liquor, or acid, is made up of sulfurous acid and a bisulfite of one of the four above bases. The sulfurous acid is usually produced by burning sulfur or pyrites and absorbing the SO_2 in liquor. Normally, part of the sulfurous acid is converted to the base bisulfite to buffer the cooking action. During the cooking action, it is necessary to vent the digester occasionally as the pressure rises within the digester. These vent gases contain large quantities of sulfur dioxide and, therefore, are recovered for reuse into the cooking acid.

Upon completion of the cooking cycle the contents of the digester, consisting of cooked chips and spent liquor, are discharged into a tank. During this operation some water vapor and fumes escape to the atmosphere from the tank vent. The pulp then goes through a washing stage, where the spent liquor is separated from the fibers. The washed pulp is either shipped or kept within the plant for further processing.

The spent liquor that was washed out of the pulp can be discarded or, as an alternative, can be concentrated by evaporation and run through a recovery cycle. The concentrated liquor is sprayed into a furnace where the organic compounds are burned. The residual inorganic compounds may be collected and reused in the manufacture of cooking acid.

3.4.2 BASIC DESCRIPTION OF SULFITE FLOW DIAGRAMS

The basic assumptions which were made in the development of each sulfite flow diagram, the diagram itself, and the energy balance are presented on the following pages.

SULFITE DIAGRAM NO. 1

TYPICAL LOCATION

Western United States.

TYPICAL AGE OF EQUIPMENT

Over fifteen years.

GENERAL

Sulfite Flow Diagram No. 1 is based upon the use of magnesium acid sulfite with chemical recovery. Cooking is done in batch digesters. There are perhaps 3 to 6 mills in the U. S. which are illustrated by this flow diagram.

EMISSIONS

Because of the possible variations in this process arrangement, valid data are not available to provide ranges of emissions.

ASSUMPTIONS

The following assumptions have been made in developing the flow diagram:

A. PULP MILL

1. Pulp Yield = 43 Percent.

2. Cooking Acid: Combined SO_2 = 1.2 Percent
Free SO_2 = 7.3 Percent
Total SO_2 = 8.5 Percent

B. POWER PLANT

1. Excess Air: 30 percent for bark, 10 percent for oil and 20 percent for coal.

2. Bark burning equipment is a low set spreader stoker.

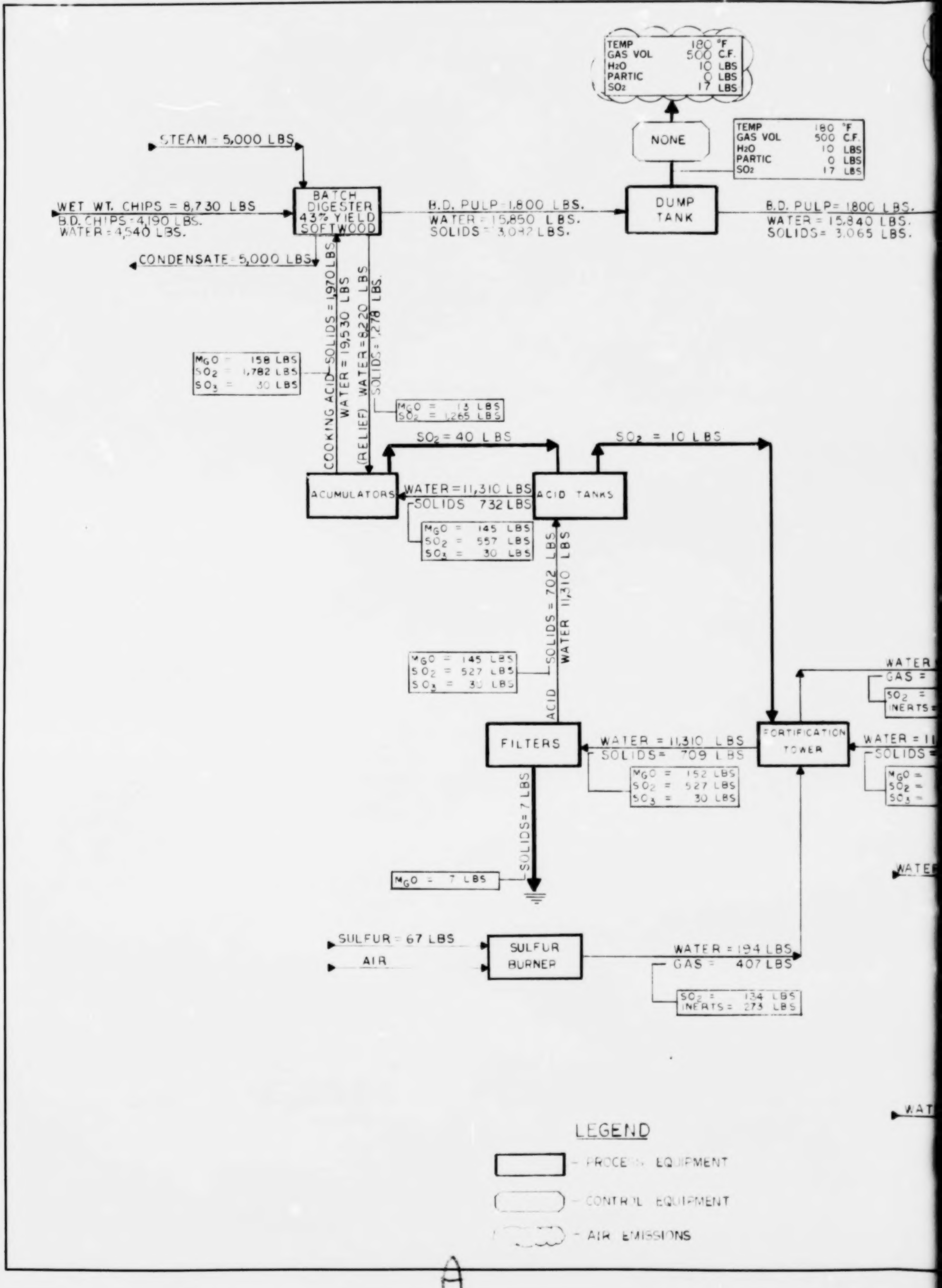
3. Reinjection from dust collector: 50 percent for bark and 0 percent for coal.

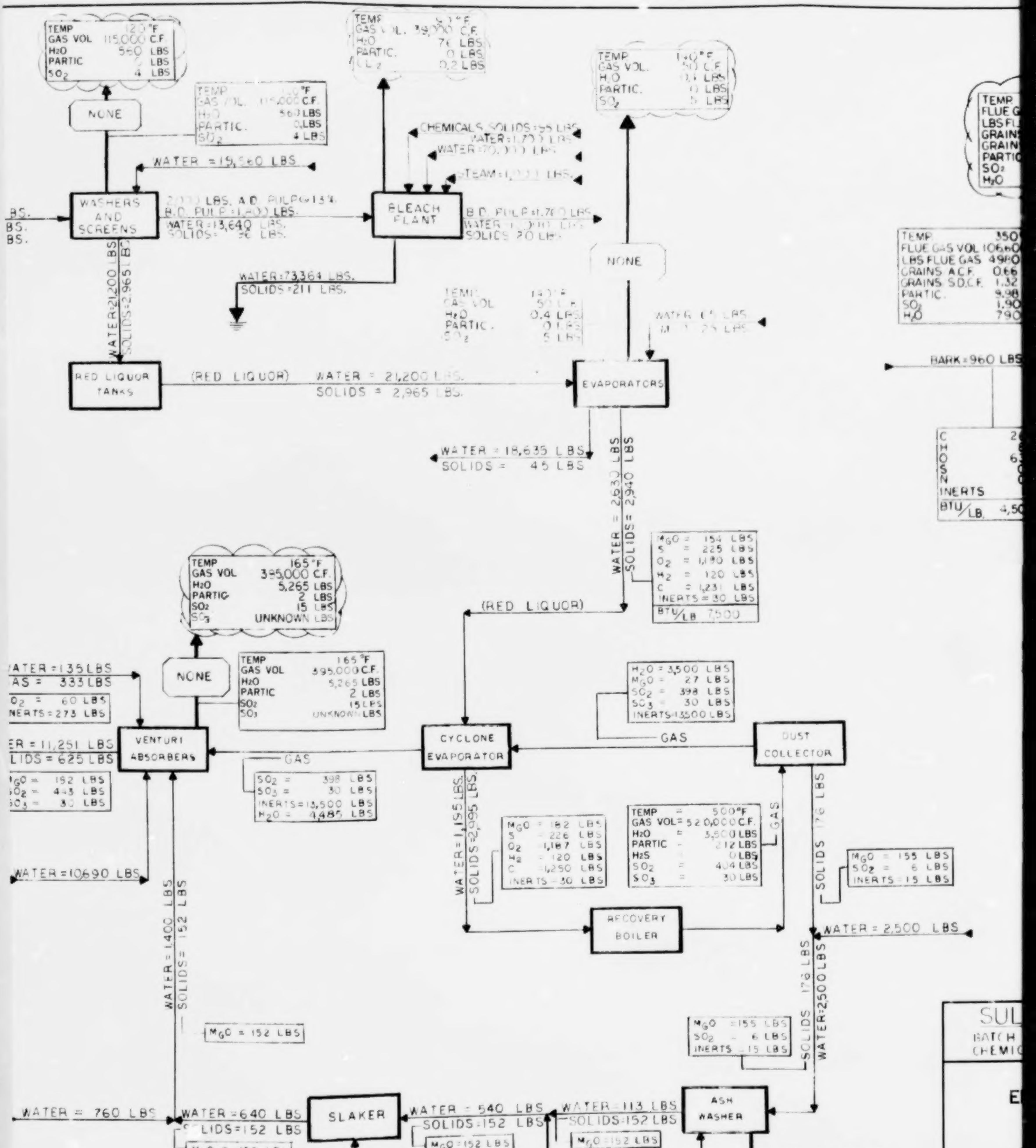
4. Design efficiencies for dust collector: 82 percent on bark and 92 percent on coal with annual operating efficiencies of 80 percent and 90 percent, respectively.

5. Bark is burned 24 hours per day at a controlled rate.

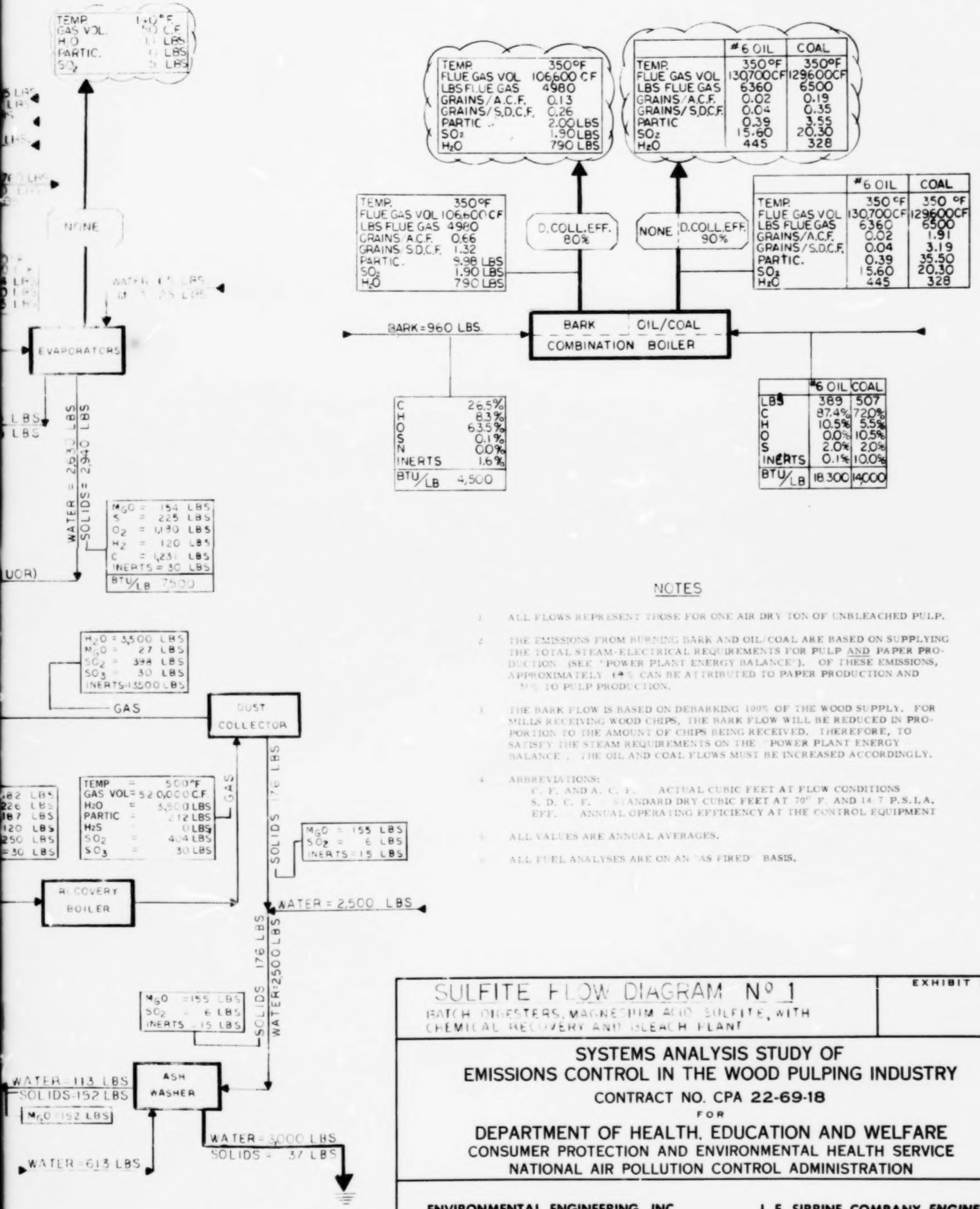
6. Unburned combustible in refuse leaving the dust collector is 40 percent.

7. Coal firing based on pulverized coal.


TEMP 180 °F
GAS VOL 500 C.F.
H2O 10 LBS
PARTIC 0 LBS
SO2 17 LBS
NONE
TEMP 180 °F
GAS VOL 500 C.F.
H2O 10 LBS
PARTIC 0 LBS
SO2 17 LBS
STEAM = 5,000 LBS.
WET WT. CHIPS = 8,730 LBS
B.D. CHIPS = 4,190 LBS.
WATER = 4,540 LBS.
BATCH DIGESTER 43% YIELD SOFTWOOD
CONDENSATE = 5,000 LBS.
B.D. PULP = 1,800 LBS.
WATER = 15,850 LBS.
SOLIDS = 3,082 LBS.
DUMP TANK
B.D. PULP = 1800 LBS.
WATER = 15,840 LBS.
SOLIDS = 3,065 LBS.
COOKING ACID SOLIDS = 1,970 LBS
WATER = 19,530 LBS
(RELIEF) WATER = 8,220 LBS
SOLIDS = 1,278 LBS.
MgO = 158 LBS
SO2 = 1,782 LBS
SO3 = 30 LBS
MgO = 13 LBS
SO2 = 1,265 LBS
SO2 = 40 LBS
SO2 = 10 LBS
ACUMULATORS
WATER = 11,310 LBS
SOLIDS 732 LBS
ACID TANKS
MgO = 145 LBS
SO2 = 557 LBS
SO3 = 30 LBS
ACID SOLIDS = 702 LBS
WATER 11,310 LBS
MgO = 145 LBS
SO2 = 527 LBS
SO3 = 30 LBS
FILTERS
WATER = 11,310 LBS
SOLIDS = 709 LBS
MgO = 152 LBS
SO2 = 527 LBS
SO3 = 30 LBS
FORTIFICATION TOWER
WATER
GAS =
SO2 =
INERTS =
WATER = 11
SOLIDS =
MgO =
SO2 =
SO3 =
WATER
SOLIDS = 7 LBS
MgO = 7 LBS
SULFUR = 67 LBS
AIR
SULFUR BURNER
WATER = 194 LBS
GAS = 407 LBS
SO2 = 134 LBS
INERTS = 273 LBS
WAT
LEGEND
PROCESS EQUIPMENT
CONTROL EQUIPMENT
AIR EMISSIONS
A


TEMP 120 °F
GAS VOL 115,000 C.F.
H2O 560 LBS
PARTIC 0 LBS
SO2 4 LBS
NONE
TEMP 120 °F
GAS VOL. 115,000 C.F.
H2O 560 LBS
PARTIC. 0 LBS
SO2 4 LBS
WATER = 19,560 LBS
WASHERS AND SCREENS
BS.
BS.
BS.
2,000 LBS. A.D. PULP@13%
B.D. PULP = 1,800 LBS.
WATER = 13,640 LBS.
SOLIDS = 36 LBS.
WATER = 21,200 LBS
SOLIDS = 2,965 LBS
RED LIQUOR TANKS
(RED LIQUOR)
WATER = 21,200 LBS.
SOLIDS = 2,965 LBS.
TEMP 60 °F
GAS VOL. 39,000 C.F.
H2O 76 LBS
PARTIC. 0 LBS
CL2 0.2 LBS
CHEMICALS SOLIDS = 55 LBS
WATER = 1,700 LBS
WATER = 70,000 LBS
STEAM = 1,000 LBS.
BLEACH PLANT
B.D. PULP = 1,760 LBS
WATER = 1,000 LBS
SOLIDS 20 LBS
WATER = 73,364 LBS.
SOLIDS = 211 LBS.
TEMP 140 °F
GAS VOL 50 C.F.
H2O 0.4 LBS
PARTIC. 0 LBS
SO2 5 LBS
TEMP 140 °F
GAS VOL. 50 C.F.
H2O 0.4 LBS
PARTIC. 0 LBS
SO2 5 LBS
NONE
WATER 65 LBS
MGO 25 LBS
EVAPORATORS
WATER = 18,635 LBS
SOLIDS = 45 LBS
WATER = 2,630 LBS
SOLIDS = 2,940 LBS
MGO = 154 LBS
S = 225 LBS
O2 = 1,180 LBS
H2 = 120 LBS
C = 1,231 LBS
INERTS = 30 LBS
BTU/LB 7,500
(RED LIQUOR)
TEMP
FLUE G
LBS FL
GRAINS
GRAIN
PARTIC
SO2
H2O
TEMP 350
FLUE GAS VOL 106,60
LBS FLUE GAS 4980
GRAINS A.C.F. 0.66
GRAINS S.D.C.F. 1.32
PARTIC. 9.98
SO2 1.90
H2O 790
BARK = 960 LBS
C 26
H
O 63
S 0
N 0
INERTS
BTU/LB. 4,50
TEMP 165 °F
GAS VOL 395,000 C.F.
H2O 5,265 LBS
PARTIC 2 LBS
SO2 15 LBS
SO3 UNKNOWN LBS
NONE
TEMP 165 °F
GAS VOL 395,000 C.F.
H2O 5,265 LBS
PARTIC 2 LBS
SO2 15 LBS
SO3 UNKNOWN LBS
WATER = 135 LBS
GAS = 333 LBS
O2 = 60 LBS
INERTS = 273 LBS
ER = 11,251 LBS
LIDS = 625 LBS
MGO = 152 LBS
SO2 = 443 LBS
SO3 = 30 LBS
VENTURI ABSORBERS
GAS
SO2 = 398 LBS
SO3 = 30 LBS
INERTS = 13,500 LBS
H2O = 4,485 LBS
WATER = 10690 LBS
WATER = 1,400 LBS
SOLIDS = 152 LBS
MGO = 152 LBS
CYCLONE EVAPORATOR
H2O = 3,500 LBS
MGO = 27 LBS
SO2 = 398 LBS
SO3 = 30 LBS
INERTS = 13,500 LBS
GAS
DUST COLLECTOR
WATER = 1,155 LBS.
SOLIDS = 2,995 LBS.
MGO = 182 LBS
S = 226 LBS
O2 = 1,187 LBS
H2 = 120 LBS
C = 1,250 LBS
INERTS = 30 LBS
TEMP = 500 °F
GAS VOL = 520,000 C.F.
H2O = 3,500 LBS
PARTIC = 212 LBS
H2S = 0 LBS
SO2 = 404 LBS
SO3 = 30 LBS
GAS
RECOVERY BOILER
SOLIDS 176 LBS
MGO = 155 LBS
SO2 = 6 LBS
INERTS = 15 LBS
WATER = 2,500 LBS
SOLIDS 176 LBS
WATER = 2,500 LBS
MGO = 155 LBS
SO2 = 6 LBS
INERTS = 15 LBS
WATER = 760 LBS
WATER = 640 LBS
SOLIDS = 152 LBS
MGO = 152 LBS
SLAKER
STEAM = 100 LBS
WATER = 540 LBS
SOLIDS = 152 LBS
MGO = 152 LBS
WATER = 427 LBS
WATER = 113 LBS
SOLIDS = 152 LBS
MGO = 152 LBS
WATER = 613 LBS
ASH WASHER
WATER = 3,000 LBS
SOLIDS = 37 LBS
SUL
BATCH
CHEMIC
E
C
ENVIRO

B


TEMP 1+0°F
GAS VOL. 50 C.F.
H2O 11 LBS
PARTIC. 0 LBS
SO2 5 LBS
NONE
WATER 65 LBS
EVAPORATORS
WATER = 2,630 LBS
SOLIDS = 2,940 LBS
MgO = 154 LBS
S = 225 LBS
O2 = 1,180 LBS
H2 = 120 LBS
C = 1,231 LBS
INERTS = 30 LBS
BTU/LB 7500
UOR)
TEMP. 350°F
FLUE GAS VOL 106600 C.F
LBS FLUE GAS 4980
GRAINS/A.C.F. 0.13
GRAINS/S.D.C.F. 0.26
PARTIC 2.00 LBS
SO2 1.90 LBS
H2O 790 LBS
#6 OIL / COAL
TEMP. 350°F / 350°F
FLUE GAS VOL 130,700 CF / 129600 CF
LBS FLUE GAS 6360 / 6500
GRAINS/A.C.F. 0.02 / 0.19
GRAINS/S.D.C.F. 0.04 / 0.35
PARTIC 0.39 / 3.55
SO2 15.60 / 20.30
H2O 445 / 328
TEMP. 350°F
FLUE GAS VOL 106,600 CF
LBS FLUE GAS 4980
GRAINS/A.C.F. 0.66
GRAINS/S.D.C.F. 1.32
PARTIC. 9.98 LBS
SO2 1.90 LBS
H2O 790 LBS
D.COLL.EFF. 80%
NONE
D.COLL.EFF. 90%
#6 OIL / COAL
TEMP. 350°F / 350°F
FLUE GAS VOL 130,700 CF / 129600 CF
LBS FLUE GAS 6360 / 6500
GRAINS/A.C.F. 0.02 / 1.91
GRAINS/S.D.C.F. 0.04 / 3.19
PARTIC. 0.39 / 35.50
SO2 15.60 / 20.30
H2O 445 / 328
BARK = 960 LBS.
BARK
OIL/COAL
COMBINATION BOILER
C 26.5%
H 8.3%
O 63.5%
S 0.1%
N 0.0%
INERTS 1.6%
BTU/LB 4,500
#6 OIL / COAL
LBS 389 / 507
C 87.4% / 72.0%
H 10.5% / 5.5%
O 0.0% / 10.5%
S 2.0% / 2.0%
INERTS 0.1% / 10.0%
BTU/LB 18300 / 14000
H2O = 3,500 LBS
MgO = 27 LBS
SO2 = 398 LBS
SO3 = 30 LBS
INERTS = 13,500 LBS
GAS
DUST COLLECTOR
TEMP = 500°F
GAS VOL = 520,000 C.F.
H2O = 3,500 LBS
PARTIC = 212 LBS
H2S = 0 LBS
SO2 = 404 LBS
SO3 = 30 LBS
GAS
SOLIDS 176 LBS
MgO = 155 LBS
SO2 = 6 LBS
INERTS = 15 LBS
RECOVERY BOILER
WATER = 2,500 LBS
SOLIDS 176 LBS
WATER = 2,500 LBS
MgO = 155 LBS
SO2 = 6 LBS
INERTS = 15 LBS
WATER = 113 LBS
SOLIDS = 152 LBS
MgO = 152 LBS
ASH WASHER
WATER = 613 LBS
WATER = 3,000 LBS
SOLIDS = 37 LBS

NOTES

1. ALL FLOWS REPRESENT THOSE FOR ONE AIR DRY TON OF UNBLEACHED PULP.
2. THE EMISSIONS FROM BURNING BARK AND OIL/COAL ARE BASED ON SUPPLYING THE TOTAL STEAM-ELECTRICAL REQUIREMENTS FOR PULP AND PAPER PRODUCTION (SEE "POWER PLANT ENERGY BALANCE"). OF THESE EMISSIONS, APPROXIMATELY 10% CAN BE ATTRIBUTED TO PAPER PRODUCTION AND 90% TO PULP PRODUCTION.
3. THE BARK FLOW IS BASED ON DEBARKING 100% OF THE WOOD SUPPLY. FOR MILLS RECEIVING WOOD CHIPS, THE BARK FLOW WILL BE REDUCED IN PROPORTION TO THE AMOUNT OF CHIPS BEING RECEIVED. THEREFORE, TO SATISFY THE STEAM REQUIREMENTS ON THE "POWER PLANT ENERGY BALANCE", THE OIL AND COAL FLOWS MUST BE INCREASED ACCORDINGLY.
4. ABBREVIATIONS:
   - C. F. AND A. C. F. = ACTUAL CUBIC FEET AT FLOW CONDITIONS
   - S. D. C. F. = STANDARD DRY CUBIC FEET AT 70° F. AND 14.7 P.S.I.A.
   - EFF. = ANNUAL OPERATING EFFICIENCY AT THE CONTROL EQUIPMENT
5. ALL VALUES ARE ANNUAL AVERAGES.
6. ALL FUEL ANALYSES ARE ON AN "AS FIRED" BASIS.

SULFITE FLOW DIAGRAM Nº 1

BATCH DIGESTERS, MAGNESIUM ACID SULFITE, WITH CHEMICAL RECOVERY AND BLEACH PLANT

EXHIBIT NO.

SYSTEMS ANALYSIS STUDY OF EMISSIONS CONTROL IN THE WOOD PULPING INDUSTRY

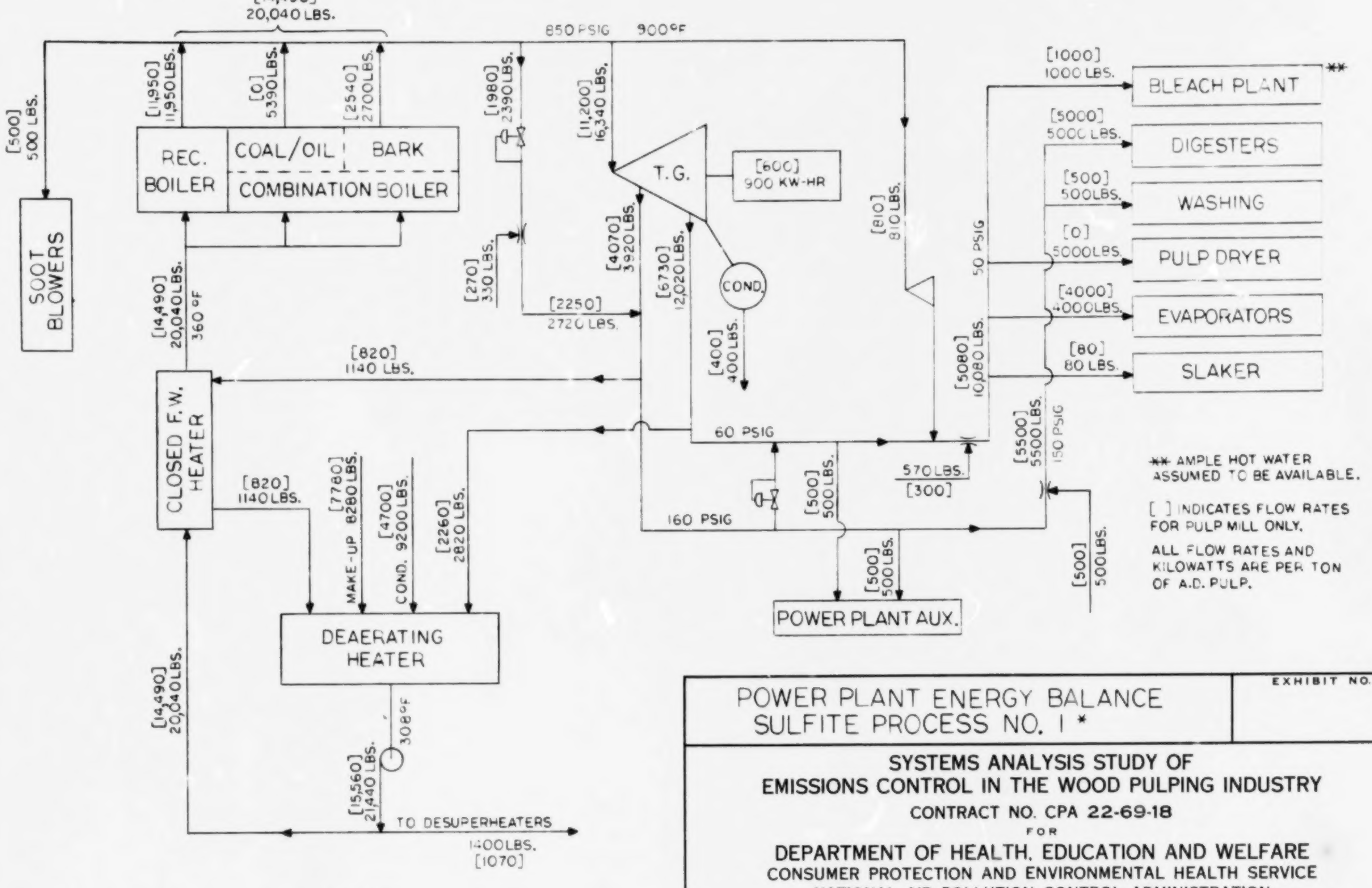
CONTRACT NO. CPA 22-69-18

FOR

DEPARTMENT OF HEALTH, EDUCATION AND WELFARE
CONSUMER PROTECTION AND ENVIRONMENTAL HEALTH SERVICE
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

ENVIRONMENTAL ENGINEERING, INC.
GAINESVILLE, FLORIDA

J. E. SIRRINE COMPANY, ENGINEERS
GREENVILLE, S. C.


[14,490]
20,040 LBS.
850 PSIG 900°F
[500]
500 LBS.
SOOT BLOWERS
[11,950]
11,950 LBS.
[0]
5390 LBS.
[2540]
2700 LBS.
REC. BOILER
COAL/OIL
BARK
COMBINATION BOILER
[1980]
2390 LBS.
[270]
330 LBS.
[2250]
2720 LBS.
[11,200]
16,340 LBS.
T. G.
[600]
900 KW-HR
[4070]
3920 LBS.
[6730]
12,020 LBS.
COND.
[400]
400 LBS.
[810]
810 LBS.
[14,490]
20,040 LBS.
360 °F
[820]
1140 LBS.
CLOSED F.W. HEATER
[820]
1140 LBS.
[7780]
MAKE-UP 8280 LBS.
[4700]
COND. 9200 LBS.
[2260]
2820 LBS.
DEAERATING HEATER
308°F
[15,560]
21,440 LBS.
[14,490]
20,040 LBS.
TO DESUPERHEATERS
1400 LBS.
[1070]
60 PSIG
160 PSIG
[500]
500 LBS.
[500]
500 LBS.
POWER PLANT AUX.
570 LBS.
[300]
[5080]
10,080 LBS.
50 PSIG
[5500]
5500 LBS.
150 PSIG
[500]
500 LBS.
[1000]
1000 LBS.
BLEACH PLANT **
[5000]
5000 LBS.
DIGESTERS
[500]
500 LBS.
WASHING
[0]
5000 LBS.
PULP DRYER
[4000]
4000 LBS.
EVAPORATORS
[80]
80 LBS.
SLAKER

** AMPLE HOT WATER ASSUMED TO BE AVAILABLE.

[ ] INDICATES FLOW RATES FOR PULP MILL ONLY.

ALL FLOW RATES AND KILOWATTS ARE PER TON OF A.D. PULP.

* BATCH DIGESTERS, MAGNESIUM ACID SULFITE WITH CHEMICAL RECOVERY AND BLEACH PLANT.

POWER PLANT ENERGY BALANCE
SULFITE PROCESS NO. 1 *

EXHIBIT NO.

SYSTEMS ANALYSIS STUDY OF
EMISSIONS CONTROL IN THE WOOD PULPING INDUSTRY
CONTRACT NO. CPA 22-69-18
FOR
DEPARTMENT OF HEALTH, EDUCATION AND WELFARE
CONSUMER PROTECTION AND ENVIRONMENTAL HEALTH SERVICE
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

ENVIRONMENTAL ENGINEERING, INC.
GAINESVILLE, FLORIDA

J. E. SIRRINE COMPANY, *ENGINEERS*
GREENVILLE, S. C.

SULFITE DIAGRAM NO. 2

TYPICAL LOCATION

Western United States.

TYPICAL AGE OF EQUIPMENT

Over 15 years.

GENERAL

Sulfite Flow Diagram No. 2 is based on the use of calcium acid sulfite, without chemical recovery. Cooking is done in batch digesters. There are perhaps 15 to 25 mills in the U. S. which are illustrated by this flow diagram.

EMISSIONS

Because of the possible variations in this process arrangement, valid data are not available to provide ranges of emissions.

ASSUMPTIONS

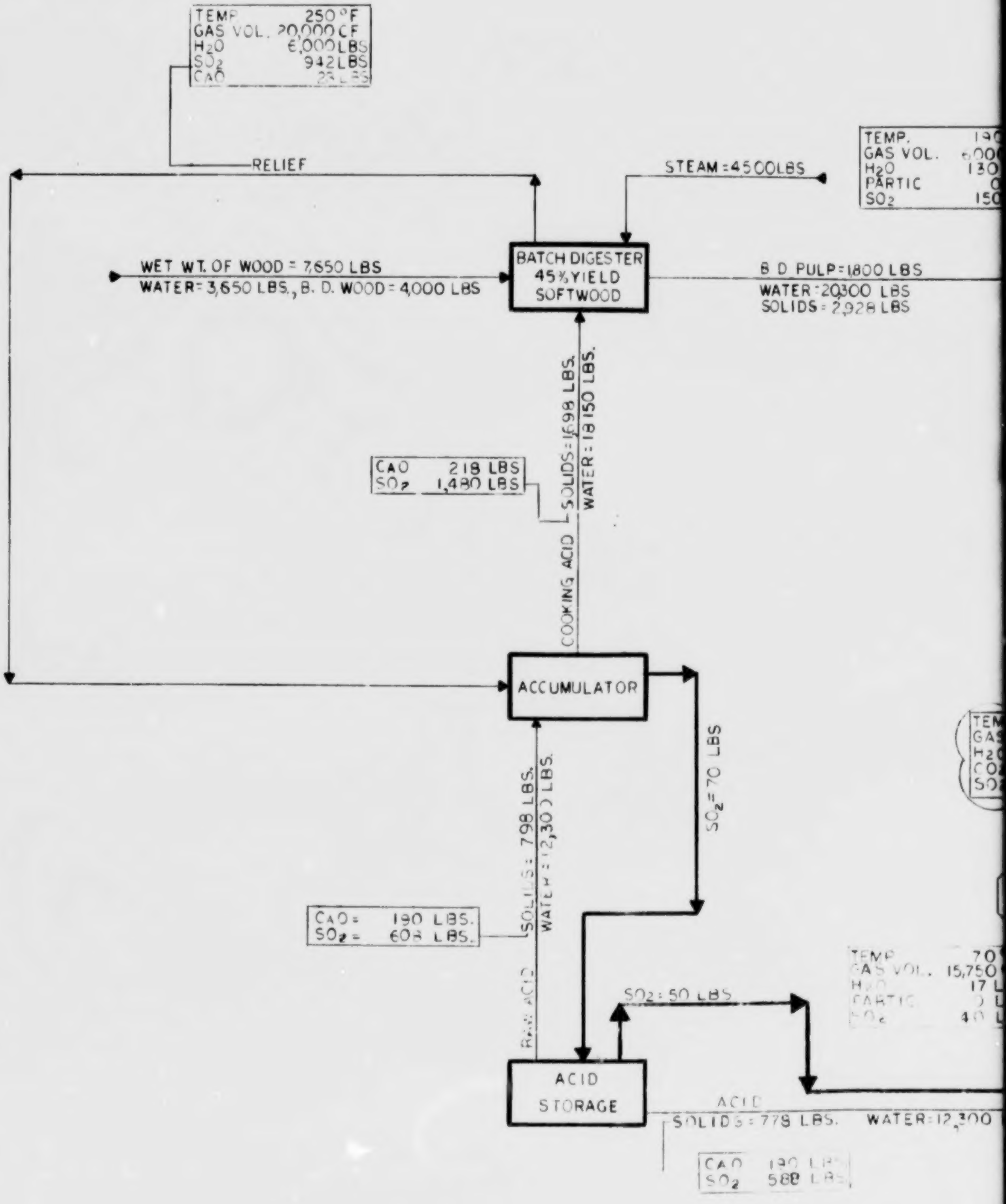
The following assumptions have been made in developing the flow diagrams:

A. PULP MILL

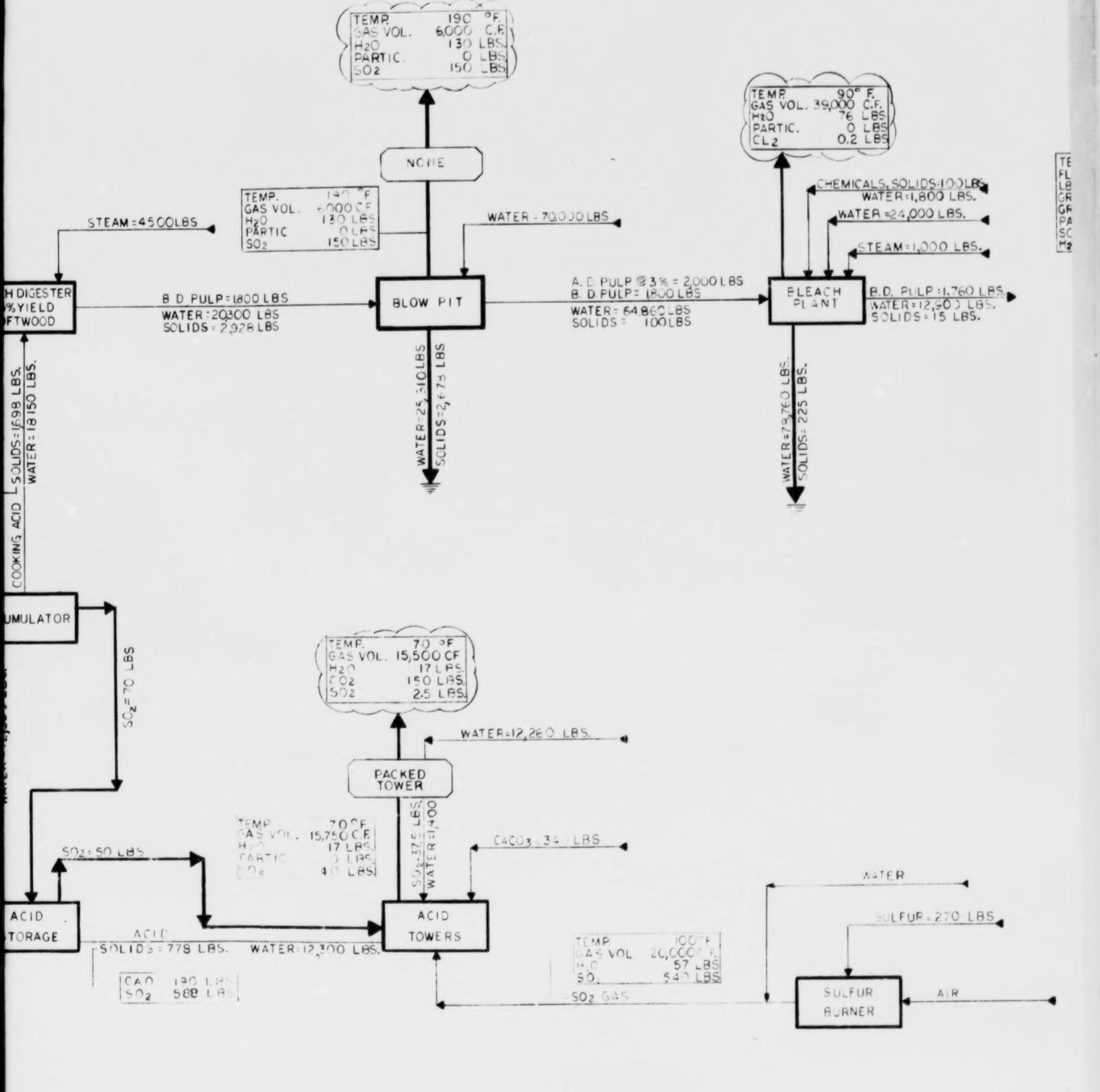
1. Pulp Yield = 45 Percent
2. Cooking Acid: Combined SO_2 = 1.25 Percent
   Free SO_2 = 7.00 Percent
   Total SO_2 = 8.25 Percent

B. POWER PLANT

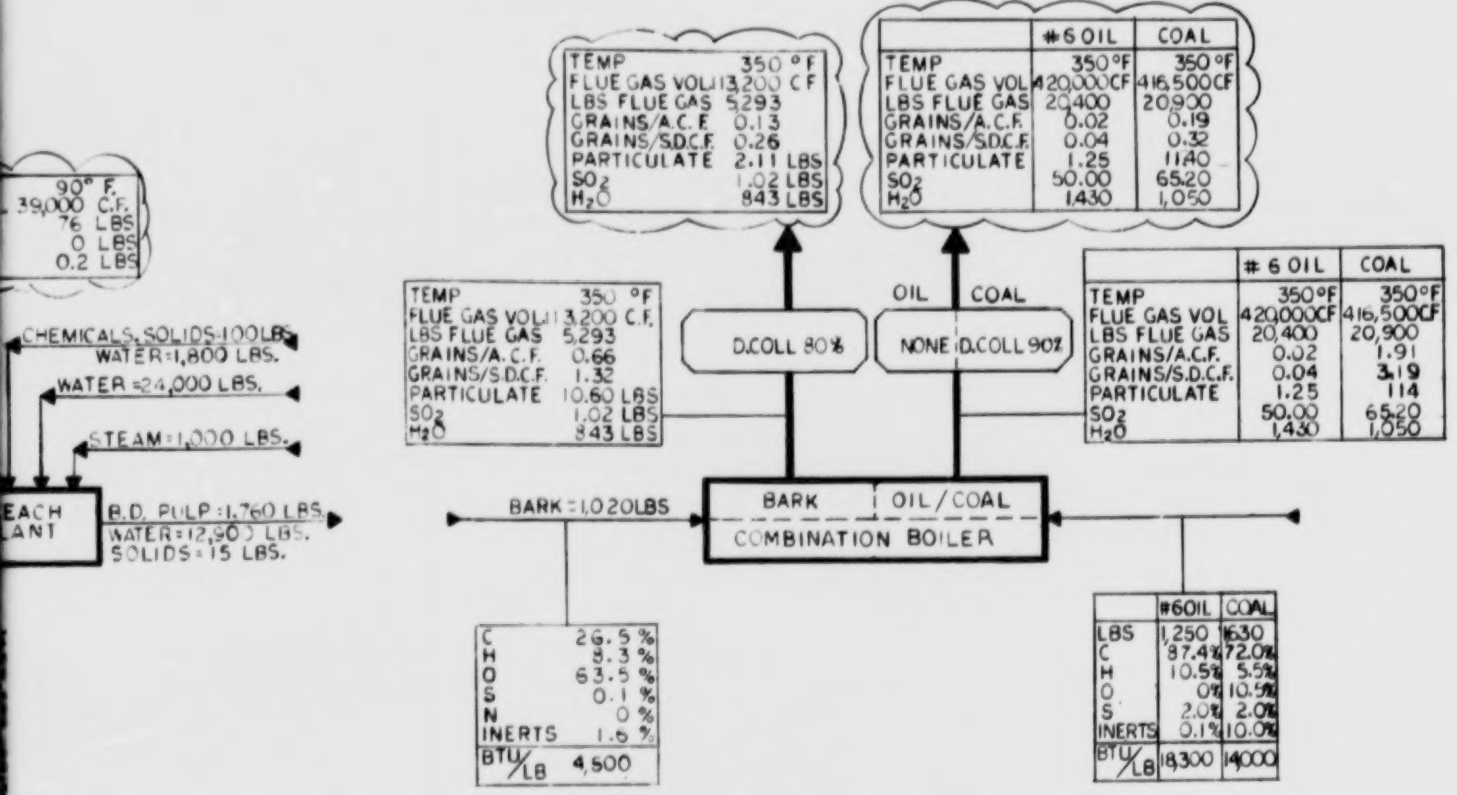
1. Excess air: 30 percent for bark, 10 percent for oil and 20 percent for coal.
2. Bark burning equipment is a low set spreader stoker.
3. Reinjection from dust collector: 50 percent for bark and 0 percent for coal.
4. Design efficiencies for dust collector: 82 percent on bark and 92 percent on coal with annual operating efficiencies of 80 percent and 90 percent, respectively.
5. Bark is burned 24 hours per day at a controlled rate.
6. Unburned combustible in refuse leaving the dust collector is 40 percent.
7. Coal firing based on pulverized coal.


TEMP 250 °F
GAS VOL. 20,000 CF
H2O 6,000 LBS
SO2 942 LBS
CAO 23 LBS
RELIEF
STEAM = 4500 LBS
TEMP. 190
GAS VOL. 6,000
H2O 130
PARTIC 0
SO2 150
WET WT. OF WOOD = 7,650 LBS
WATER = 3,650 LBS., B. D. WOOD = 4,000 LBS
BATCH DIGESTER
45% YIELD
SOFTWOOD
B D PULP = 1,800 LBS
WATER = 20,300 LBS
SOLIDS = 2,928 LBS
CAO 218 LBS
SO2 1,480 LBS
SOLIDS = 1,698 LBS.
WATER = 18150 LBS.
COOKING ACID
ACCUMULATOR
SO2 = 70 LBS
CAO = 190 LBS.
SO2 = 608 LBS.
SOLIDS = 798 LBS.
WATER = 12,300 LBS.
RAW ACID
TEMP 70
GAS VOL. 15,750
H2O 17 L
PARTIC 0 L
SO2 40 L
SO2 = 50 LBS
ACID
STORAGE
ACID
SOLIDS = 778 LBS.
WATER = 12,300
CAO 190 LBS
SO2 588 LBS

A


TEMP. 190 °F.
GAS VOL. 6,000 C.F.
H2O 130 LBS.
PARTIC. 0 LBS.
SO2 150 LBS.
NONE
TEMP. 190 °F
GAS VOL. 6,000 C.F.
H2O 130 LBS
PARTIC 0 LBS
SO2 150 LBS
STEAM = 4500 LBS
WATER = 70,000 LBS
H DIGESTER
%YIELD
FTWOOD
B D PULP = 1800 LBS
WATER: 20,300 LBS
SOLIDS = 2,928 LBS
BLOW PIT
WATER = 25,310 LBS
SOLIDS = 2,478 LBS
SOLIDS = 1,698 LBS.
WATER: 18 150 LBS.
COOKING ACID
TEMP. 90° F.
GAS VOL. 39,000 C.F.
H2O 76 LBS
PARTIC. 0 LBS
CL2 0.2 LBS
CHEMICALS. SOLIDS = 100 LBS.
WATER = 1,800 LBS.
WATER = 24,000 LBS.
STEAM = 1,000 LBS.
A. D. PULP @ 3% = 2,000 LBS
B. D. PULP = 1800 LBS
WATER = 64,860 LBS
SOLIDS = 100 LBS
BLEACH PLANT
B.D. PULP = 1,760 LBS.
WATER = 12,900 LBS.
SOLIDS = 15 LBS.
WATER = 79,760 LBS.
SOLIDS = 225 LBS.
TE
FL
LB
GR
GR
PA
SC
H2
UMULATOR
SO2 = 70 LBS
TEMP. 70 °F
GAS VOL. 15,500 CF
H2O 17 LBS.
CO2 150 LBS.
SO2 2.5 LBS.
WATER = 12,260 LBS.
PACKED TOWER
TEMP. 70 °F
GAS VOL. 15,750 C.F.
H2O 17 LBS.
PARTIC 0 LBS.
SO2 40 LBS.
SO2 = 375 LBS.
WATER = 1,400
CACO3 = 34 LBS
SO2 = 50 LBS
ACID
STORAGE
ACID
SOLIDS = 778 LBS.
WATER = 12,300 LBS.
ACID TOWERS
CAO 190 LBS
SO2 588 LBS
WATER
SULFUR = 270 LBS
TEMP. 100 °F
GAS VOL. 20,000 C.F.
H2O 57 LBS
SO2 540 LBS
SO2 GAS
SULFUR BURNER
AIR
LEGEND:
— PROC
— CONT
— AIR

B


90° F.
39,000 C.F.
76 LBS
0 LBS
0.2 LBS
CHEMICALS, SOLIDS-100LBS
WATER=1,800 LBS.
WATER =24,000 LBS.
STEAM=1,000 LBS.
EACH
ANT
B.D. PULP=1,760 LBS.
WATER=12,900 LBS.
SOLIDS=15 LBS.
BARK=1,020LBS
BARK
OIL/COAL
COMBINATION BOILER
D.COLL 80%
OIL
COAL
NONE
D.COLL 90%

TEMP	350 °F
FLUE GAS VOL.	13,200 C F
LBS FLUE GAS	5,293
GRAINS/A.C.F.	0.13
GRAINS/S.D.C.F.	0.26
PARTICULATE	2.11 LBS
SO_2	1.02 LBS
H_2O	843 LBS

	#6 OIL	COAL
TEMP	350°F	350°F
FLUE GAS VOL	420,000CF	416,500CF
LBS FLUE GAS	20,400	20,900
GRAINS/A.C.F.	0.02	0.19
GRAINS/S.D.C.F.	0.04	0.32
PARTICULATE	1.25	11.40
SO_2	50.00	65.20
H_2O	1,430	1,050

TEMP	350 °F
FLUE GAS VOL.	13,200 C.F.
LBS FLUE GAS	5,293
GRAINS/A.C.F.	0.66
GRAINS/S.D.C.F.	1.32
PARTICULATE	10.60 LBS
SO_2	1.02 LBS
H_2O	843 LBS

	#6 OIL	COAL
TEMP	350°F	350°F
FLUE GAS VOL	420,000CF	416,500CF
LBS FLUE GAS	20,400	20,900
GRAINS/A.C.F.	0.02	1.91
GRAINS/S.D.C.F.	0.04	3.19
PARTICULATE	1.25	114
SO_2	50.00	65.20
H_2O	1,430	1,050

C	26.5 %
H	8.3 %
O	63.5 %
S	0.1 %
N	0 %
INERTS	1.6 %
BTU/LB	4,500

	#6OIL	COAL
LBS	1,250	1630
C	87.4%	72.0%
H	10.5%	5.5%
O	0%	10.5%
S	2.0%	2.0%
INERTS	0.1%	10.0%
BTU/LB	18,300	14000

NOTES

1. ALL FLOWS REPRESENT THOSE FOR ONE AIR DRY TON OF UNBLEACHED PULP.

2. THE EMISSIONS FROM BURNING BARK AND OIL/COAL ARE BASED ON SUPPLYING THE TOTAL STEAM-ELECTRICAL REQUIREMENTS FOR PULP AND PAPER PRODUCTION (SEE "POWER PLANT ENERGY BALANCE"). OF THESE EMISSIONS, APPROXIMATELY 35 % CAN BE ATTRIBUTED TO PAPER PRODUCTION AND 65% TO PULP PRODUCTION.

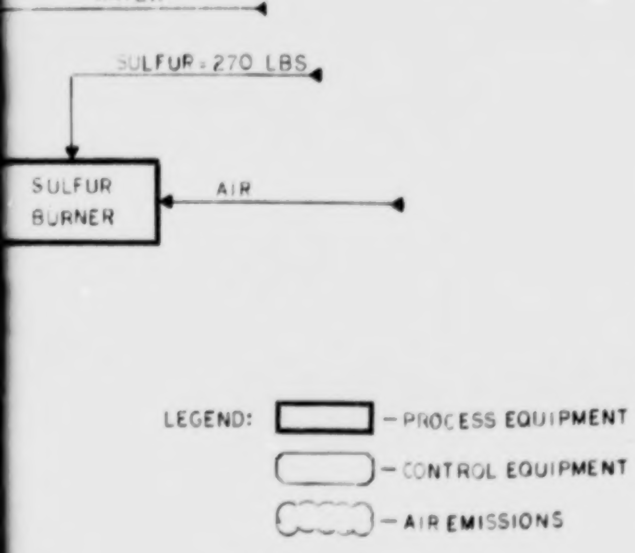
3. THE BARK FLOW IS BASED ON DEBARKING 100% OF THE WOOD SUPPLY. FOR MILLS RECEIVING WOOD CHIPS, THE BARK FLOW WILL BE REDUCED IN PROPORTION TO THE AMOUNT OF CHIPS BEING RECEIVED. THEREFORE, TO SATISFY THE STEAM REQUIREMENTS ON THE "POWER PLANT ENERGY BALANCE", THE OIL AND COAL FLOWS MUST BE INCREASED ACCORDINGLY.

4. ABBREVIATIONS

   C. F. AND A. C. F. - ACTUAL CUBIC FEET AT FLOW CONDITIONS
   S. D. C. F. - STANDARD DRY CUBIC FEET AT 70° F. AND 14.7 P.S.I.A.
   EFF. - ANNUAL OPERATING EFFICIENCY AT THE CONTROL EQUIPMENT

5. ALL VALUES ARE ANNUAL AVERAGES.

6. ALL FUEL ANALYSES ARE ON AN "AS FIRED" BASIS.


WATER
SULFUR=270 LBS
SULFUR BURNER
AIR

LEGEND:
▭ - PROCESS EQUIPMENT
▢ - CONTROL EQUIPMENT
☁ - AIR EMISSIONS

SULFITE FLOW DIAGRAM Nº 2

BATCH DIGESTERS, CALCIUM ACID SULFITE, WITH BLEACH PLANT AND WITHOUT CHEMICAL RECOVERY

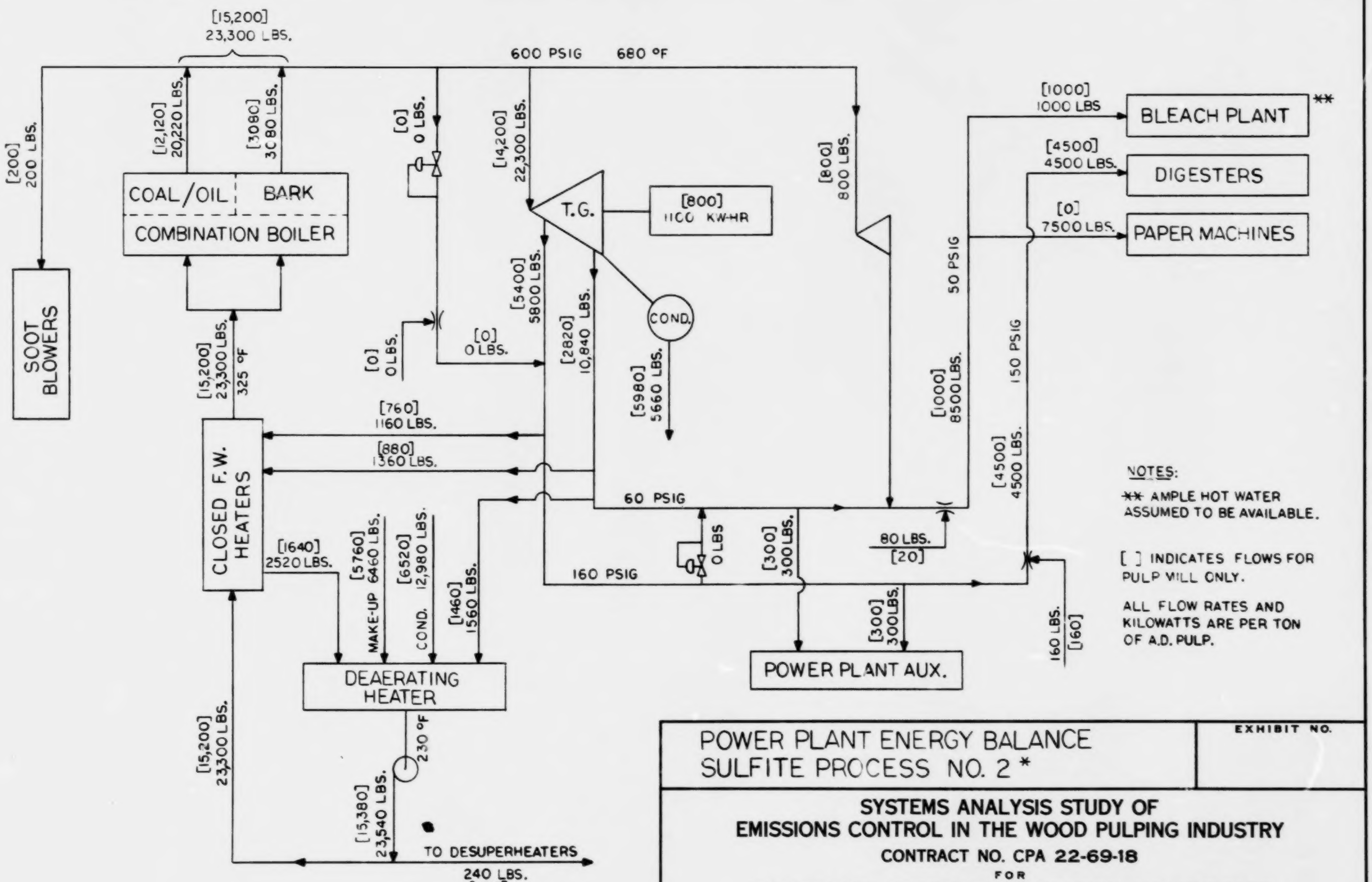
EXHIBIT NO.

SYSTEMS ANALYSIS STUDY OF EMISSIONS CONTROL IN THE WOOD PULPING INDUSTRY

CONTRACT NO. CPA 22-69-18
FOR
DEPARTMENT OF HEALTH, EDUCATION AND WELFARE
CONSUMER PROTECTION AND ENVIRONMENTAL HEALTH SERVICE
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

ENVIRONMENTAL ENGINEERING, INC.
GAINESVILLE, FLORIDA

J. E. SIRRINE COMPANY, ENGINEERS
GREENVILLE, S. C.


[15,200]
23,300 LBS.
600 PSIG
680 °F
[200]
200 LBS.
[12,120]
20,220 LBS.
[3080]
3080 LBS.
COAL/OIL
BARK
COMBINATION BOILER
SOOT BLOWERS
[0]
0 LBS.
[14,200]
22,300 LBS.
T.G.
[800]
1100 KW-HR
[800]
800 LBS.
[1000]
1000 LBS
BLEACH PLANT **
[4500]
4500 LBS.
DIGESTERS
[0]
7500 LBS.
PAPER MACHINES
50 PSIG
150 PSIG
[5400]
5800 LBS.
[2820]
10,840 LBS.
COND.
[5980]
5660 LBS.
[0]
0 LBS.
[0]
0 LBS.
[15,200]
23,300 LBS.
325 °F
[760]
1160 LBS.
[880]
1360 LBS.
[1000]
8500 LBS.
[4500]
4500 LBS.
CLOSED F.W. HEATERS
60 PSIG
0 LBS
[300]
300 LBS.
80 LBS.
[20]
[1640]
2520 LBS.
MAKE-UP
[5760]
6460 LBS.
COND.
[6520]
12,980 LBS.
[1460]
1560 LBS.
160 PSIG
[300]
300 LBS.
160 LBS.
[160]
POWER PLANT AUX.
DEAERATING HEATER
230 °F
[15,200]
23,300 LBS.
[15,380]
23,540 LBS.
TO DESUPERHEATERS
240 LBS.
[180]

NOTES:

** AMPLE HOT WATER ASSUMED TO BE AVAILABLE.

[ ] INDICATES FLOWS FOR PULP MILL ONLY.

ALL FLOW RATES AND KILOWATTS ARE PER TON OF A.D. PULP.

* BATCH DIGESTERS, CALCIUM ACID SULFITE WITHOUT CHEMICAL RECOVERY BLEACH PLANT.

POWER PLANT ENERGY BALANCE SULFITE PROCESS NO. 2 *

EXHIBIT NO.

SYSTEMS ANALYSIS STUDY OF EMISSIONS CONTROL IN THE WOOD PULPING INDUSTRY

CONTRACT NO. CPA 22-69-18

FOR

DEPARTMENT OF HEALTH, EDUCATION AND WELFARE
CONSUMER PROTECTION AND ENVIRONMENTAL HEALTH SERVICE
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

ENVIRONMENTAL ENGINEERING, INC.
GAINESVILLE, FLORIDA

J. E. SIRRINE COMPANY, *ENGINEERS*
GREENVILLE, S. C.

SULFITE DIAGRAM NO. 3

TYPICAL LOCATION

Western United States

TYPICAL AGE OF EQUIPMENT

Converted to magnesium base within the past five years.

GENERAL

Sulfite Flow Diagram No. 3 is based on the use of magnesium bisulfite liquor (Magnefite), with chemical recovery. Cooking is done in batch digesters. There are perhaps 2 to 4 mills in the U. S. which are illustrated by this flow diagram.

EMISSIONS

Because of the possible variations in the process arrangement, valid data are not available to provide ranges of emissions.

ASSUMPTIONS

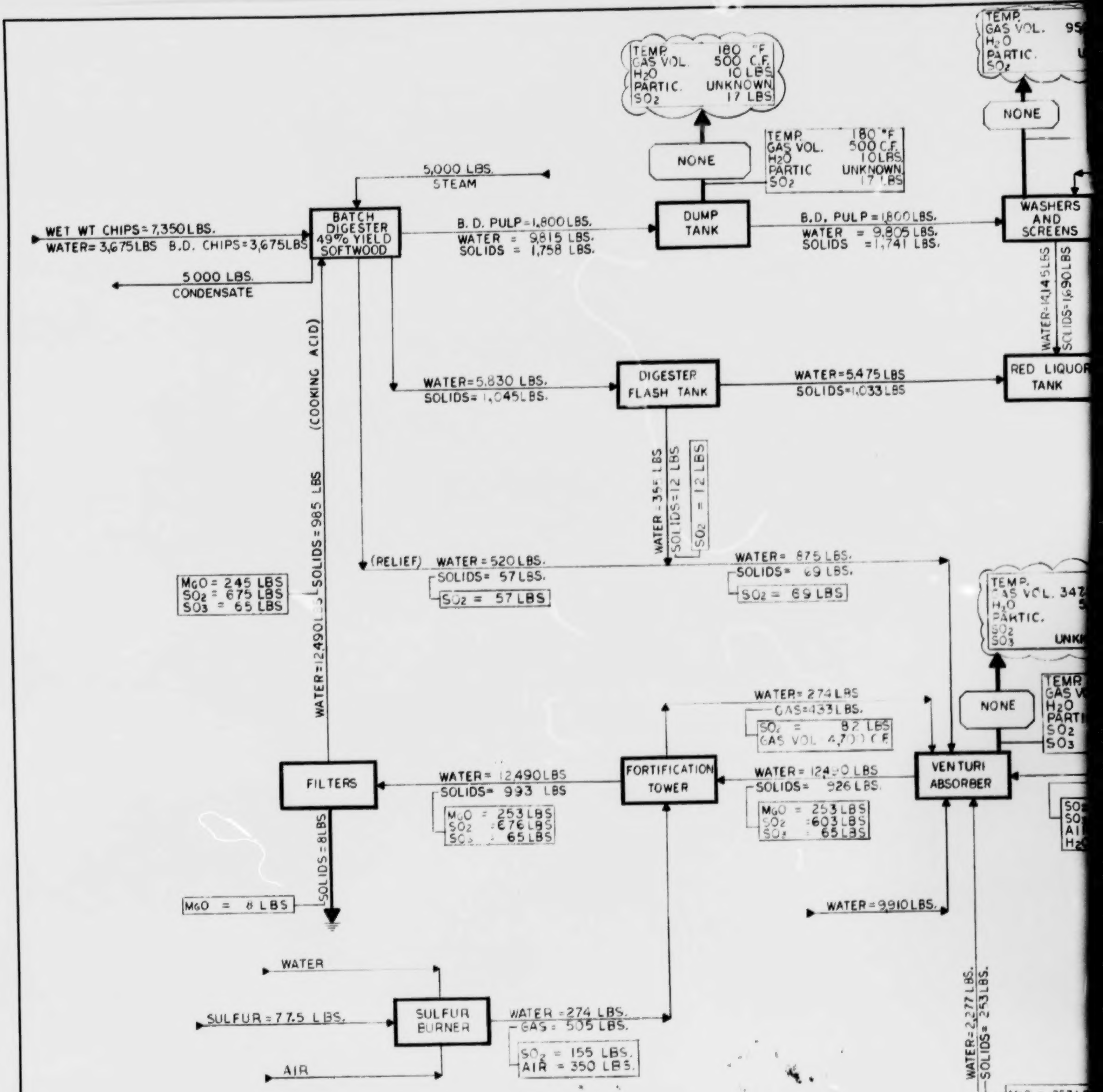
The following assumptions have been made in developing the flow diagram:

A. PULP MILL

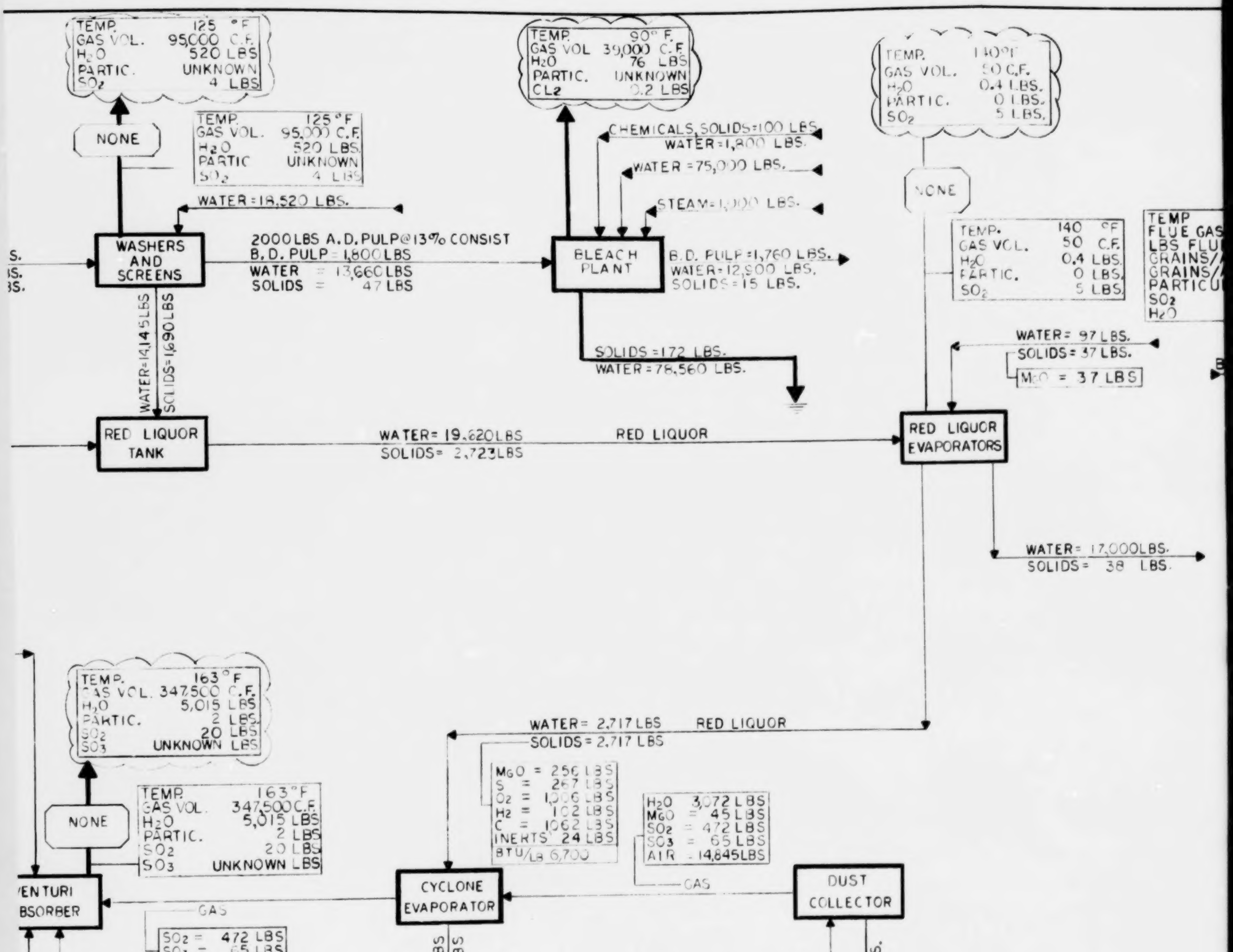
1. Pulp Yield = 50 Percent
2. Cooking Liquor:
   - Combined SO_2 = 2.5 Percent.
   - Free SO_2 = 2.5 Percent
   - Total SO_2 = 5.0 Percent

B. POWER PLANT

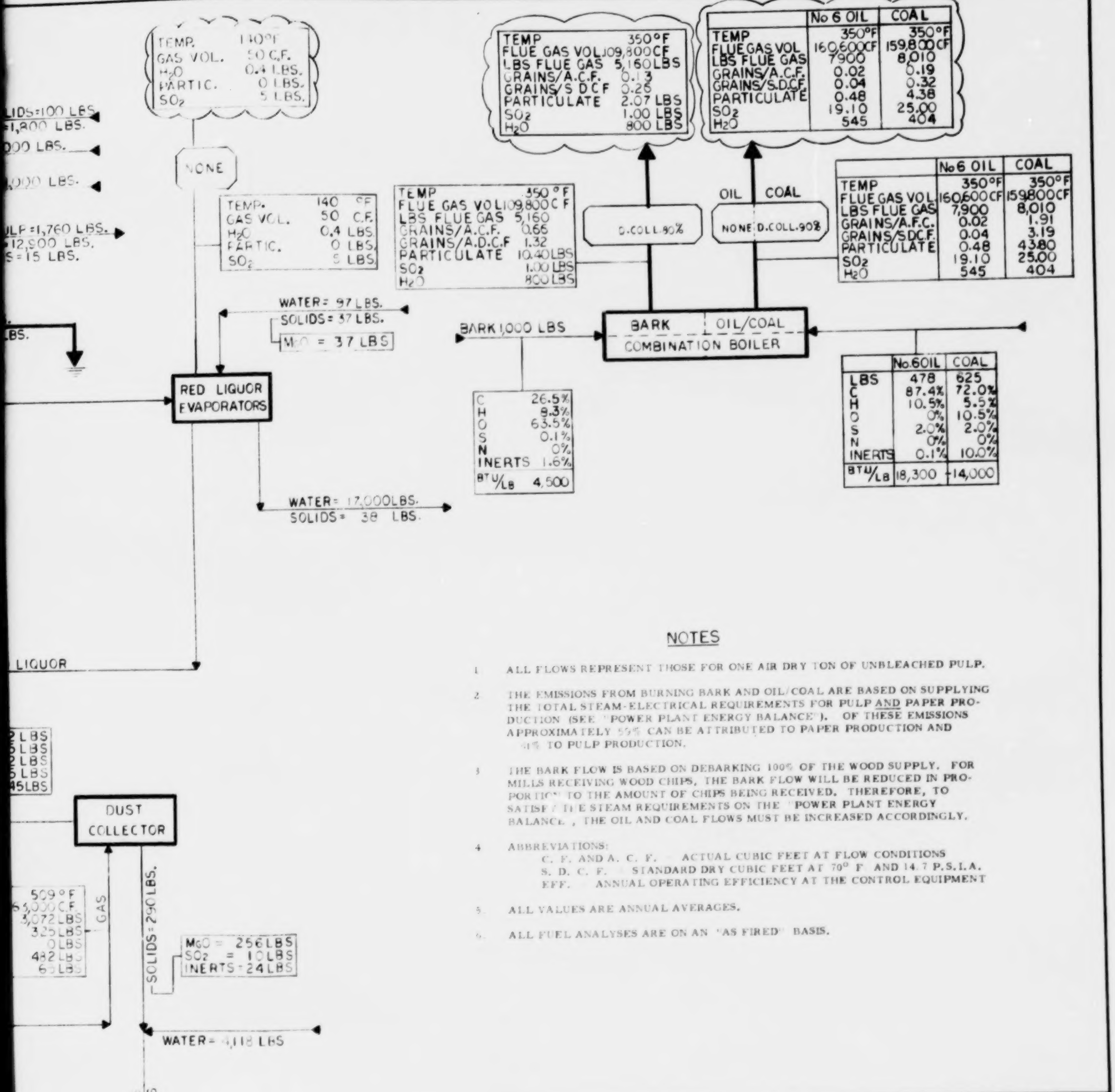
1. Excess Air: 30 percent for bark, 10 percent for oil and 20 percent for coal.
2. Bark burning equipment is a low set spreader stoker.
3. Reinjection from dust collector: 50 percent for bark and 0 percent for coal.
4. Design efficiencies for dust collector: 82 percent on bark and 92 percent on coal with annual operating efficiencies of 80 percent and 90 percent, respectively.
5. Bark is burned 24 hours per day at a controlled rate.
6. Unburned combustible in refuse leaving the dust collector is 40 percent.
7. Coal firing based on pulverized coal.


TEMP. 180 °F
GAS VOL. 500 C.F.
H2O 10 LBS
PARTIC. UNKNOWN
SO2 17 LBS
NONE
TEMP. 180 °F
GAS VOL. 500 C.F.
H2O 10 LBS
PARTIC UNKNOWN
SO2 17 LBS
TEMP.
GAS VOL. 95
H2O
PARTIC. U
SO2
NONE
5,000 LBS.
STEAM
WET WT CHIPS = 7,350 LBS.
WATER = 3,675 LBS B.D. CHIPS = 3,675 LBS
BATCH DIGESTER 49% YIELD SOFTWOOD
B. D. PULP = 1,800 LBS.
WATER = 9,815 LBS.
SOLIDS = 1,758 LBS.
DUMP TANK
B.D. PULP = 1,800 LBS.
WATER = 9,805 LBS.
SOLIDS = 1,741 LBS.
WASHERS AND SCREENS
WATER = 14,145 LBS
SOLIDS = 1,690 LBS
5000 LBS.
CONDENSATE
(COOKING ACID)
WATER = 5,830 LBS.
SOLIDS = 1,045 LBS.
DIGESTER FLASH TANK
WATER = 5,475 LBS
SOLIDS = 1,033 LBS
RED LIQUOR TANK
WATER = 355 LBS
SOLIDS = 12 LBS
SO2 = 12 LBS
WATER = 12,490 LBS
SOLIDS = 985 LBS
MgO = 245 LBS
SO2 = 675 LBS
SO3 = 65 LBS
(RELIEF) WATER = 520 LBS.
SOLIDS = 57 LBS.
SO2 = 57 LBS
WATER = 875 LBS.
SOLIDS = 69 LBS.
SO2 = 69 LBS
TEMP.
GAS VOL. 347
H2O 5
PARTIC.
SO2
SO3 UNK
WATER = 274 LBS
GAS = 433 LBS.
SO2 = 82 LBS
GAS VOL. 4,700 C.F.
NONE
TEMP.
GAS V
H2O
PARTI
SO2
SO3
FILTERS
WATER = 12,490 LBS
SOLIDS = 993 LBS
MgO = 253 LBS
SO2 = 676 LBS
SO3 = 65 LBS
FORTIFICATION TOWER
WATER = 12,490 LBS
SOLIDS = 926 LBS.
MgO = 253 LBS
SO2 = 603 LBS
SO3 = 65 LBS
VENTURI ABSORBER
SO
SO
AI
H2
SOLIDS = 8 LBS
MgO = 8 LBS
WATER = 9,910 LBS.
WATER
SULFUR = 77.5 LBS.
SULFUR BURNER
WATER = 274 LBS.
GAS = 505 LBS.
SO2 = 155 LBS.
AIR = 350 LBS.
AIR
WATER = 2,277 LBS.
SOLIDS = 253 LBS.
MgO = 253 LB
WATER = 1270 LBS
WATER = 1,007
SOLIDS = 253
ST
LEGEND
- PROCESS EQUIPMENT
- CONTROL EQUIPMENT
- AIR EMISSIONS

A


TEMP. 125 °F
GAS VOL. 95,000 C.F.
H_2O 520 LBS
PARTIC. UNKNOWN
SO_2 4 LBS
NONE
TEMP. 125 °F
GAS VOL. 95,000 C.F.
H_2O 520 LBS.
PARTIC UNKNOWN
SO_2 4 LBS
WATER = 18,520 LBS.
WASHERS AND SCREENS
2000 LBS A.D. PULP @ 13% CONSIST
B. D. PULP = 1,800 LBS
WATER = 13,660 LBS
SOLIDS = 47 LBS
WATER = 14,145 LBS
SOLIDS = 1,690 LBS
RED LIQUOR TANK
WATER = 19,620 LBS
SOLIDS = 2,723 LBS
RED LIQUOR
TEMP. 90° F.
GAS VOL 39,000 C.F.
H_2O 76 LBS
PARTIC. UNKNOWN
CL_2 0.2 LBS
CHEMICALS, SOLIDS = 100 LBS
WATER = 1,800 LBS.
WATER = 75,000 LBS.
STEAM = 1,000 LBS.
BLEACH PLANT
B. D. PULP = 1,760 LBS.
WATER = 12,800 LBS.
SOLIDS = 15 LBS.
SOLIDS = 172 LBS.
WATER = 78,560 LBS.
TEMP. 140 °F
GAS VOL. 50 C.F.
H_2O 0.4 LBS.
PARTIC. 0 LBS.
SO_2 5 LBS.
NONE
TEMP. 140 °F
GAS VOL. 50 C.F.
H_2O 0.4 LBS.
PARTIC. 0 LBS.
SO_2 5 LBS.
TEMP
FLUE GAS
LBS FLU
GRAINS/
GRAINS/
PARTICU
SO_2
H_2O
WATER = 97 LBS.
SOLIDS = 37 LBS.
MGO = 37 LBS
RED LIQUOR EVAPORATORS
WATER = 17,000 LBS.
SOLIDS = 38 LBS.
TEMP. 163 °F
GAS VOL. 347,500 C.F.
H_2O 5,015 LBS
PARTIC. 2 LBS
SO_2 20 LBS
SO_3 UNKNOWN LBS
NONE
TEMP. 163 °F
GAS VOL. 347,500 C.F.
H_2O 5,015 LBS
PARTIC. 2 LBS
SO_2 20 LBS
SO_3 UNKNOWN LBS
VENTURI ABSORBER
GAS
SO_2 = 472 LBS
SO_3 = 65 LBS
AIR = 14845 LBS
H_2O = 4,166 LBS
WATER = 2,717 LBS
SOLIDS = 2,717 LBS
RED LIQUOR
MGO = 256 LBS
S = 267 LBS
O_2 = 1,006 LBS
H_2 = 102 LBS
C = 1062 LBS
INERTS = 24 LBS
BTU/LB 6,700
H_2O 3,072 LBS
MGO = 45 LBS
SO_2 = 472 LBS
SO_3 = 65 LBS
AIR = 14,845 LBS
GAS
CYCLONE EVAPORATOR
DUST COLLECTOR
WATER = 1623 LBS
SOLIDS = 2,762 LBS
MGO = 301 LBS
S = 267 LBS
O_2 = 1,006 LBS
H_2 = 102 LBS
C = 1,062 LBS
INERTS = 24 LBS
TEMP 509 °F
GAS VOL 453,000 C.F.
H_2O 3,072 LBS
PARTIC. 325 LBS
H_2S 0 LBS
SO_2 482 LBS
SO_3 65 LBS
GAS
SOLIDS = 290 LBS.
MGO = 256 LBS
SO_2 = 10 LBS
INERTS = 24 LBS
RECOVERY BOILER
WATER = 4,118 LBS
WATER = 2,277 LBS.
SOLIDS = 253 LBS.
MGO = 253 LBS
WATER = 4,118 LBS
SOLIDS = 290 LBS
MGO 256 LBS
SO_2 = 10 LBS
INERTS = 24 LBS
WATER = 1,007 LBS
SOLIDS = 253 LBS
SLAKER
WATER = 907 LBS
SOLIDS = 253 LBS
MGO = 253 LBS
WATER = 187 LBS
SOLIDS = 253 LBS.
MGO = 253 LBS
ASH WASHER
WATER = 4,471 LBS
SOLIDS = 37 LBS.
STEAM = 100 LBS
WATER = 720 LBS
WATER = 540 LBS
BA
W
ENV

B


TEMP. 140°F
GAS VOL. 50 C.F.
H2O 0.4 LBS.
PARTIC. 0 LBS.
SO2 5 LBS.
LIDS=100 LBS
=1,800 LBS.
00 LBS.
000 LBS.
ULF=1,760 LBS.
=12,900 LBS.
S=15 LBS.
BS.
NONE
TEMP. 140°F
GAS VOL. 50 C.F.
H2O 0.4 LBS.
PARTIC. 0 LBS.
SO2 5 LBS.
D.COLL.80%
OIL
COAL
NONE
D.COLL.90%
WATER=97 LBS.
SOLIDS=37 LBS.
MgO = 37 LBS
BARK 1,000 LBS
BARK
OIL/COAL
COMBINATION BOILER
RED LIQUOR EVAPORATORS
WATER=17,000 LBS.
SOLIDS=38 LBS.
LIQUOR
2 LBS
5 LBS
2 LBS
5 LBS
45 LBS
DUST COLLECTOR
509°F
63,000 C.F.
3,072 LBS
325 LBS
0 LBS
482 LBS
65 LBS
GAS
SOLIDS=290 LBS.
MgO = 256 LBS
SO2 = 10 LBS
INERTS = 24 LBS
WATER=4,118 LBS
WATER=4,118 LBS
SOLIDS=290 LBS
MgO 256 LBS
SO2 10 LBS
INERTS 24 LBS
ASH WASHER
WATER=4,471 LBS
SOLIDS=37 LBS.

TEMP	350°F
FLUE GAS VOL	109,800 CF
LBS FLUE GAS	5,160 LBS
GRAINS/A.C.F.	0.13
GRAINS/S.DCF	0.26
PARTICULATE	2.07 LBS
SO_2	1.00 LBS
H_2O	800 LBS

	No 6 OIL	COAL
TEMP	350°F	350°F
FLUE GAS VOL	160,600 CF	159,800 CF
LBS FLUE GAS	7900	8,010
GRAINS/A.C.F.	0.02	0.19
GRAINS/S.D.C.F.	0.04	0.32
PARTICULATE	0.48	4.38
SO_2	19.10	25.00
H_2O	545	404

TEMP	350°F
FLUE GAS VOL	109,800 CF
LBS FLUE GAS	5,160
GRAINS/A.C.F.	0.65
GRAINS/A.D.C.F	1.32
PARTICULATE	10.40 LBS
SO_2	1.00 LBS
H_2O	800 LBS

	No 6 OIL	COAL
TEMP	350°F	350°F
FLUE GAS VOL	160,600 CF	159,800 CF
LBS FLUE GAS	7,900	8,010
GRAINS/A.F.C.	0.02	1.91
GRAINS/S.D.C.F.	0.04	3.19
PARTICULATE	0.48	43.80
SO_2	19.10	25.00
H_2O	545	404

C	26.5%
H	9.3%
O	63.5%
S	0.1%
N	0%
INERTS	1.6%
BTU/LB	4,500

	No.6 OIL	COAL
LBS	478	625
C	87.4%	72.0%
H	10.5%	5.5%
O	0%	10.5%
S	2.0%	2.0%
N	0%	0%
INERTS	0.1%	10.0%
BTU/LB	18,300	14,000

NOTES

1. ALL FLOWS REPRESENT THOSE FOR ONE AIR DRY TON OF UNBLEACHED PULP.
2. THE EMISSIONS FROM BURNING BARK AND OIL/COAL ARE BASED ON SUPPLYING THE TOTAL STEAM-ELECTRICAL REQUIREMENTS FOR PULP AND PAPER PRODUCTION (SEE "POWER PLANT ENERGY BALANCE"). OF THESE EMISSIONS APPROXIMATELY 59% CAN BE ATTRIBUTED TO PAPER PRODUCTION AND 41% TO PULP PRODUCTION.
3. THE BARK FLOW IS BASED ON DEBARKING 100% OF THE WOOD SUPPLY. FOR MILLS RECEIVING WOOD CHIPS, THE BARK FLOW WILL BE REDUCED IN PROPORTION TO THE AMOUNT OF CHIPS BEING RECEIVED. THEREFORE, TO SATISFY THE STEAM REQUIREMENTS ON THE "POWER PLANT ENERGY BALANCE", THE OIL AND COAL FLOWS MUST BE INCREASED ACCORDINGLY.
4. ABBREVIATIONS:
   - C. F. AND A. C. F. = ACTUAL CUBIC FEET AT FLOW CONDITIONS
   - S. D. C. F. = STANDARD DRY CUBIC FEET AT 70° F AND 14.7 P.S.I.A.
   - EFF. = ANNUAL OPERATING EFFICIENCY AT THE CONTROL EQUIPMENT
5. ALL VALUES ARE ANNUAL AVERAGES.
6. ALL FUEL ANALYSES ARE ON AN "AS FIRED" BASIS.

SULFITE FLOW DIAGRAM No. 3
BATCH DIGESTERS, MAGNESIUM BISULFITE-(MAGNEFITE), WITH CHEMICAL RECOVERY AND BLEACH PLANT

EXHIBIT NO.

SYSTEMS ANALYSIS STUDY OF
EMISSIONS CONTROL IN THE WOOD PULPING INDUSTRY
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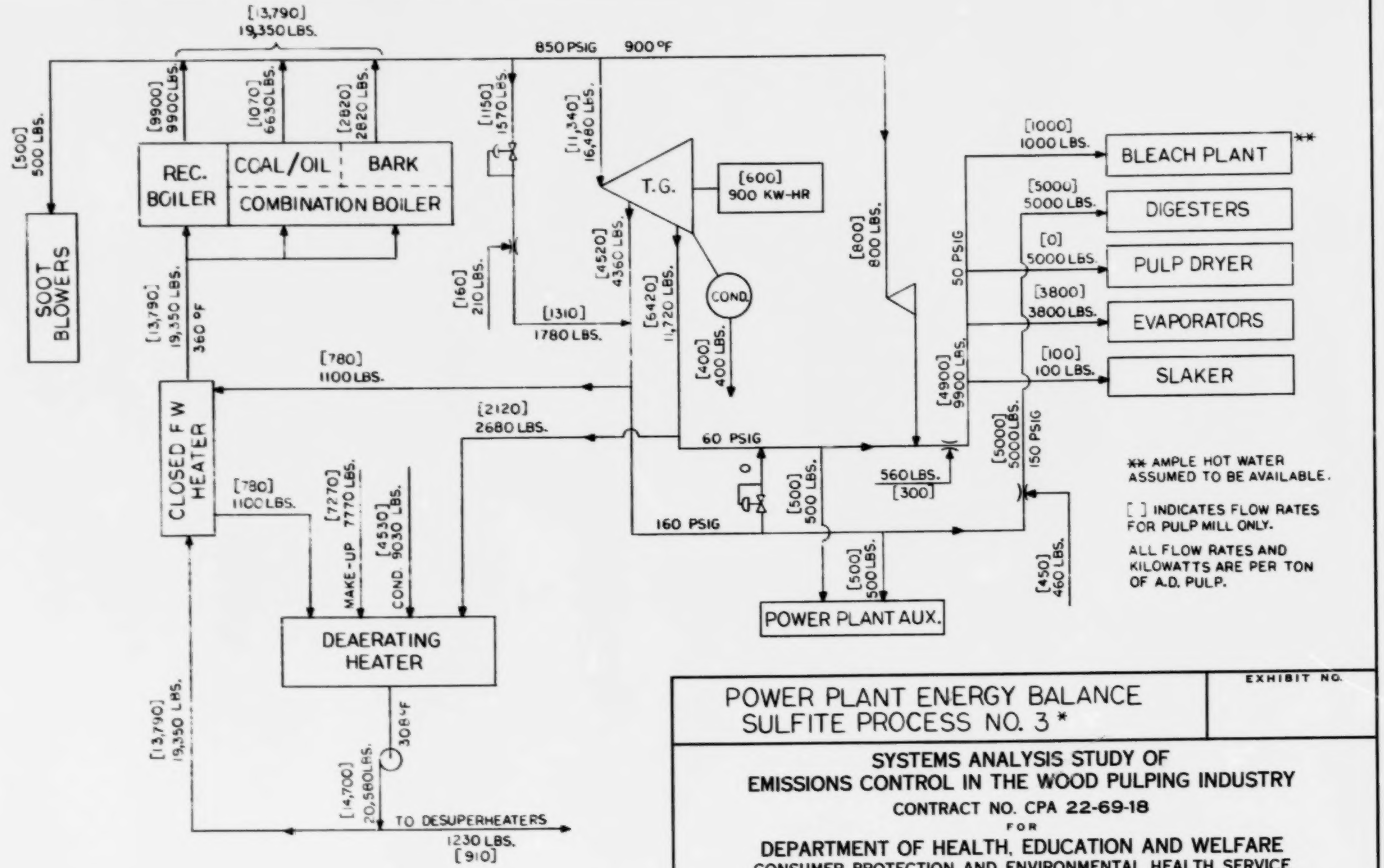
GAINESVILLE, FLORIDA

J. E. SIRRINE COMPANY, ENGINEERS
GREENVILLE, S. C.


[13,790]
19,350 LBS.
850 PSIG
900 °F
[500]
500 LBS.
SOOT BLOWERS
[9900]
9900 LBS.
[1070]
6630 LBS.
[2820]
2820 LBS.
REC. BOILER
COAL/OIL
BARK
COMBINATION BOILER
[1150]
1570 LBS.
[160]
210 LBS.
[1310]
1780 LBS.
[11,340]
16,480 LBS.
T.G.
[600]
900 KW-HR
[4520]
4360 LBS.
[6420]
11,720 LBS.
COND.
[400]
400 LBS.
[800]
800 LBS.
[13,790]
19,350 LBS.
360 °F
[780]
1100 LBS.
CLOSED F W HEATER
[2120]
2680 LBS.
60 PSIG
0
[780]
1100 LBS.
[7270]
MAKE-UP 7770 LBS.
[4530]
COND. 9030 LBS.
DEAERATING HEATER
160 PSIG
[500]
500 LBS.
[500]
500 LBS.
POWER PLANT AUX.
560 LBS.
[300]
[4900]
9900 LBS.
50 PSIG
[5000]
5000 LBS.
150 PSIG
[450]
460 LBS.
[1000]
1000 LBS.
BLEACH PLANT **
[5000]
5000 LBS.
DIGESTERS
[0]
5000 LBS.
PULP DRYER
[3800]
3800 LBS.
EVAPORATORS
[100]
100 LBS.
SLAKER
[13,790]
19,350 LBS.
308 °F
[14,700]
20,580 LBS.
TO DESUPERHEATERS
1230 LBS.
[910]

** AMPLE HOT WATER ASSUMED TO BE AVAILABLE.

[ ] INDICATES FLOW RATES FOR PULP MILL ONLY.

ALL FLOW RATES AND KILOWATTS ARE PER TON OF A.D. PULP.

* BATCH DIGESTERS, MAGNESIUM BISULFITE (MAGNEFITE), WITH CHEMICAL RECOVERY AND BLEACH PLANT.

POWER PLANT ENERGY BALANCE
SULFITE PROCESS NO. 3 *

EXHIBIT NO.

SYSTEMS ANALYSIS STUDY OF
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NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

ENVIRONMENTAL ENGINEERING, INC.
GAINESVILLE, FLORIDA

J. E. SIRRINE COMPANY, *ENGINEERS*
GREENVILLE, S. C.

SULFITE DIAGRAM NO. 4

TYPICAL LOCATION

Western United States.

TYPICAL AGE OF EQUIPMENT

Over 15 years.

GENERAL

Sulfite Flow Diagram No. 4 is based on the use of ammonium acid sulfite, with liquor incineration. Cooking is done in batch digesters. There are perhaps 1 to 4 mills in the U. S. which are illustrated by this flow diagram.

EMISSIONS

Because of the possible variations in this process arrangement, valid data are not available to provide ranges of emissions.

ASSUMPTIONS

The following assumptions have been made in developing the flow diagram:

A. PULP MILL

1. Pulp Yield = 50 Percent

2. Cooking Acid:
   - Combined SO_2 = 1.0 Percent
   - Free SO_2 = 7.0 Percent
   - Total SO_2 = 8.0 Percent

B. BLEACH PLANT

Assumed as a three stage bleach plant.

C. POWER PLANT

1. Excess air: 30 percent for bark, 10 percent for oil and 20 percent for coal.

2. Bark burning equipment is a low set spreader stoker.

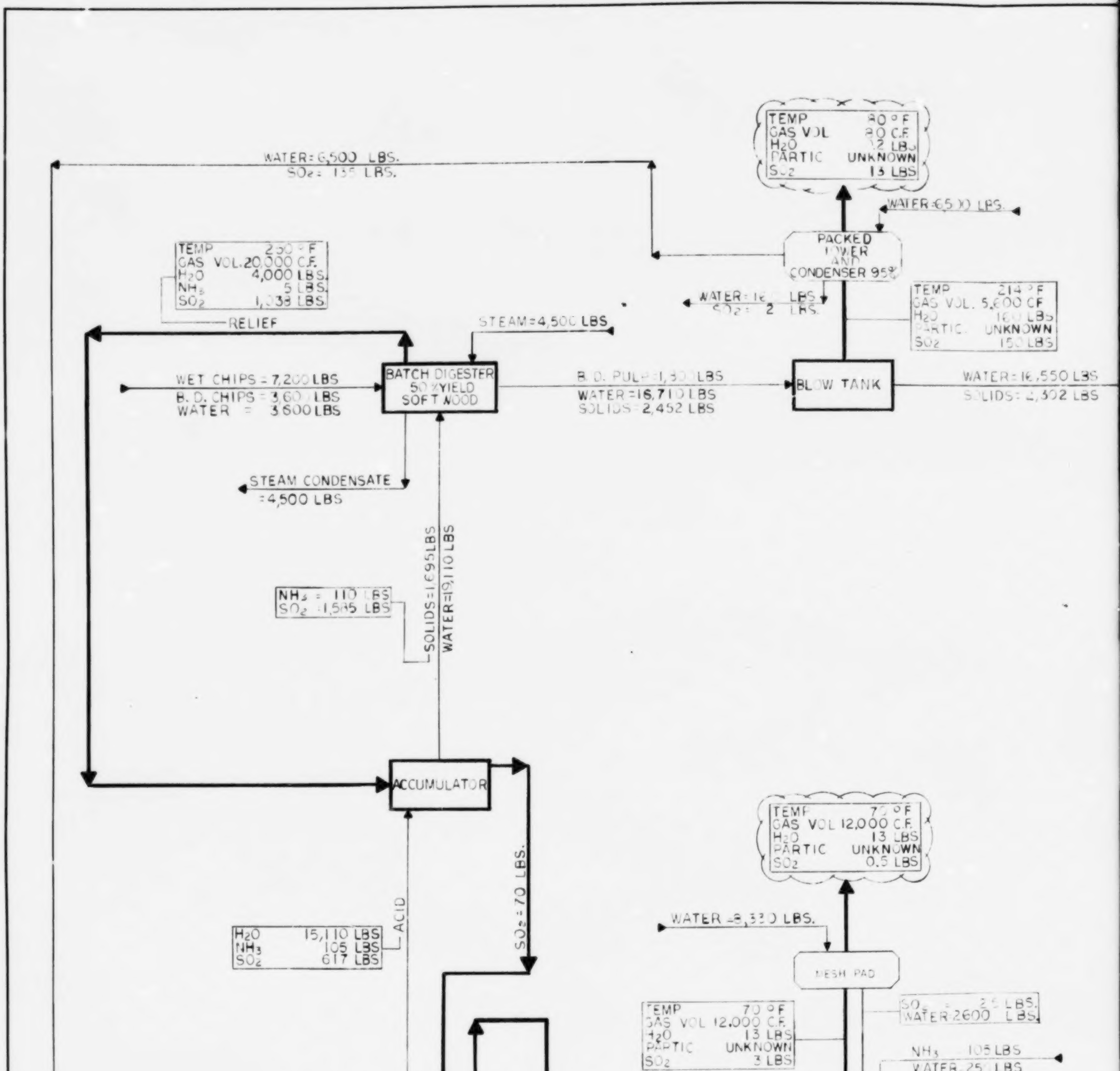
3. Reinjection from dust collector: 50 percent for bark and 0 percent for coal.

4. Design efficiencies for dust collector: 82 percent on bark and 92 percent on coal with annual operating efficiencies of 80 percent and 90 percent, respectively.

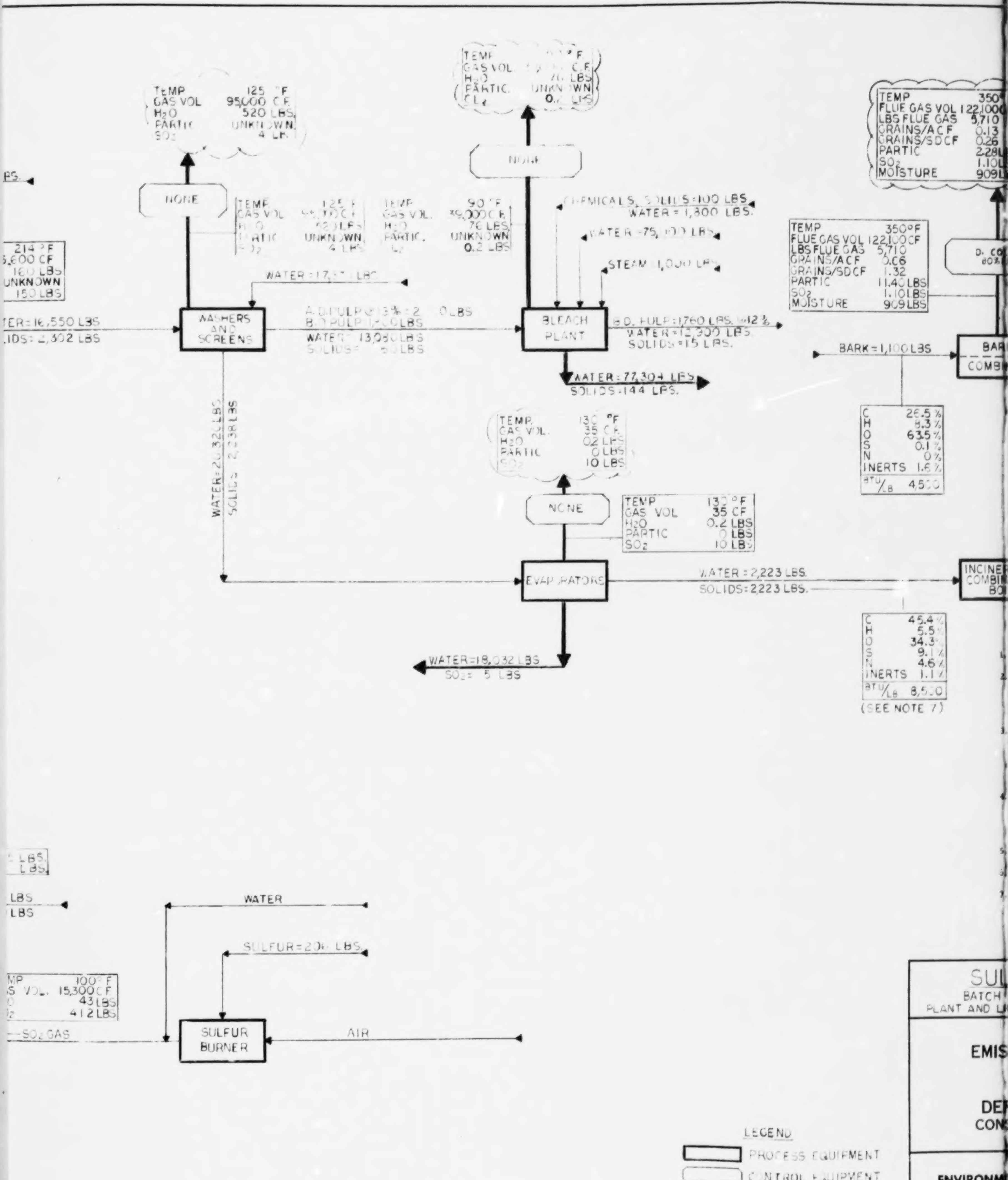
5. Bark is burned 24 hours per day at a controlled rate.

6. Unburned combustible in refuse leaving the dust collector is 40 percent.

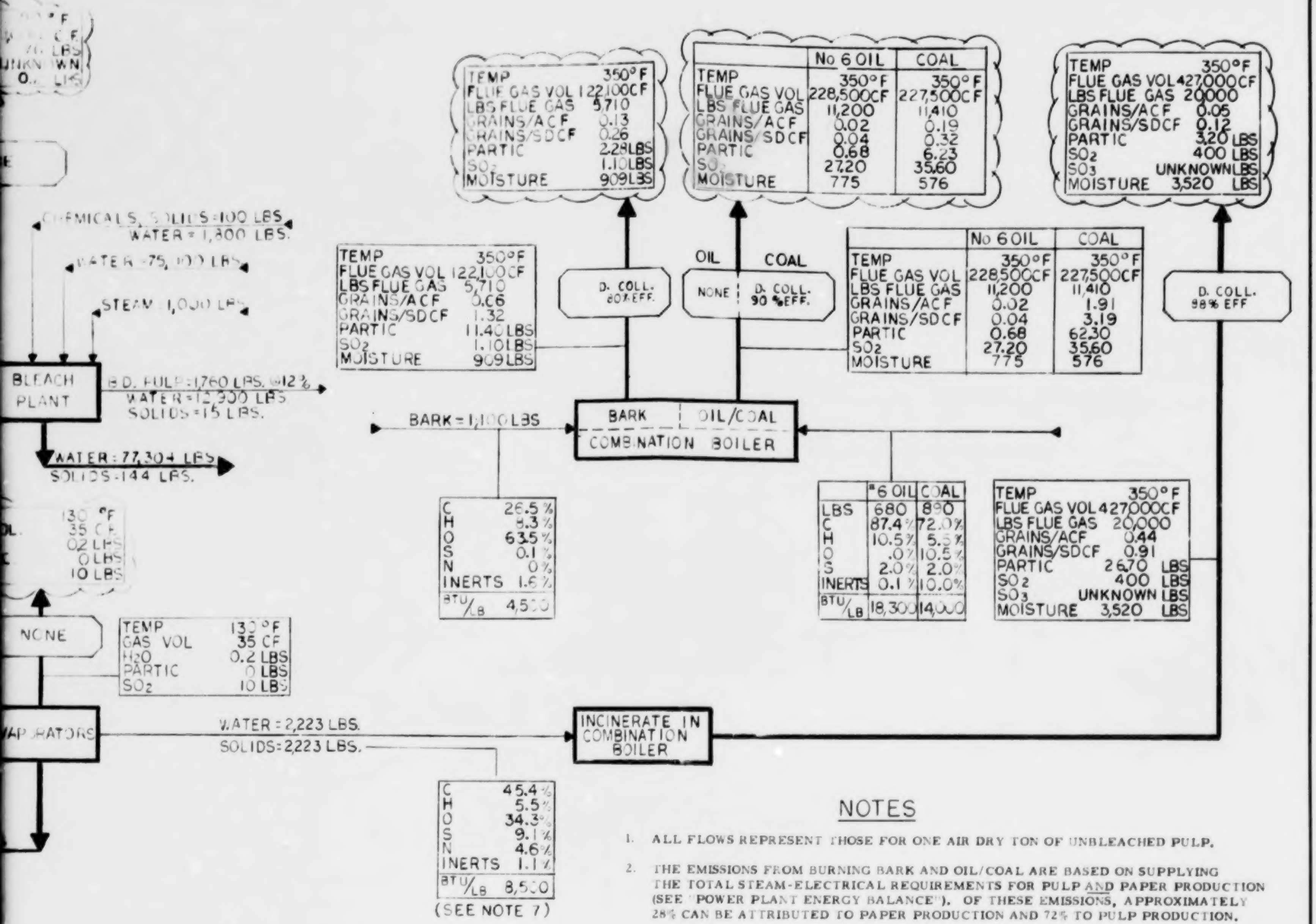
7. Coal firing based on pulverized coal.


TEMP 80 °F
GAS VOL 80 C.F.
H_2O 12 LBS
PARTIC UNKNOWN
SO_2 13 LBS
WATER = 6,500 LBS.
SO_2 = 135 LBS.
WATER = 6,500 LBS.
PACKED TOWER AND CONDENSER 95%
WATER = 160 LBS.
SO_2 = 2 LBS.
TEMP 214 °F
GAS VOL. 5,600 CF
H_2O 160 LBS
PARTIC. UNKNOWN
SO_2 150 LBS
TEMP 250 °F
GAS VOL. 20,000 C.F.
H_2O 4,000 LBS.
NH_3 5 LBS.
SO_2 1,038 LBS.
RELIEF
STEAM = 4,500 LBS
WET CHIPS = 7,200 LBS
B. D. CHIPS = 3,600 LBS
WATER = 3,600 LBS
BATCH DIGESTER 50% YIELD SOFT WOOD
B. D. PULP = 1,800 LBS
WATER = 16,710 LBS
SOLIDS = 2,452 LBS
BLOW TANK
WATER = 16,550 LBS
SOLIDS = 2,302 LBS
STEAM CONDENSATE = 4,500 LBS
NH_3 = 110 LBS
SO_2 = 1,585 LBS
SOLIDS = 1,695 LBS
WATER = 19,110 LBS
ACCUMULATOR
SO_2 = 70 LBS.
ACID
H_2O 15,110 LBS
NH_3 105 LBS
SO_2 617 LBS
ACID STORAGE
TEMP 70 °F
GAS VOL 12,000 C.F.
H_2O 13 LBS
PARTIC UNKNOWN
SO_2 0.5 LBS
WATER = 8,330 LBS.
MESH PAD
SO_2 = 2.5 LBS.
WATER 2600 LBS.
TEMP 70 °F
GAS VOL 12,000 C.F.
H_2O 13 LBS
PARTIC UNKNOWN
SO_2 3 LBS
NH_3 = 105 LBS
WATER = 250 LBS
SO_2 = 10 LBS
ABSORPTION TOWER
ACID
H_2O 8,610 LBS
NH_3 105 LBS
SO_2 422 LBS
TEMP 100 °F
GAS VOL. 15,300 C.F.
H_2O 43 LBS
SO_2 412 LBS
SO_2 GAS
WATER = 6,500 LBS
SO_2 = 60 LBS

A

TEMP 125 °F
GAS VOL 95,000 C.F.
H₂O 520 LBS.
PARTIC UNKNOWN.
SO₂ 4 LBS.
TEMP °F
GAS VOL. C.F.
H₂O 76 LBS.
PARTIC. UNKNOWN
CL₂ 0.2 LBS
NONE
NONE
TEMP 125 °F
GAS VOL 95,000 C.F.
H₂O 520 LBS.
PARTIC UNKNOWN.
SO₂ 4 LBS
TEMP 90 °F
GAS VOL. 39,000 C.F.
H₂O 76 LBS.
PARTIC. UNKNOWN
CL₂ 0.2 LBS
CHEMICALS, SOLIDS = 100 LBS.
WATER = 1,300 LBS.
WATER = 75,000 LBS.
STEAM = 1,000 LBS.
WATER = 17,5 LBS.
LBS.
214 °F
5,600 CF
160 LBS
UNKNOWN
150 LBS
TER = 16,550 LBS
LIDS = 2,302 LBS
WASHERS AND SCREENS
A.D. PULP @ 13% = 2 0 LBS
B.D. PULP 1, 0 LBS
WATER = 13,080 LBS
SOLIDS = 50 LBS
BLEACH PLANT
B.D. PULP = 1760 LBS. @12%
WATER = 12,300 LBS.
SOLIDS = 15 LBS.
WATER = 77,304 LBS.
SOLIDS = 144 LBS.
WATER = 20,320 LBS
SOLIDS = 2,238 LBS
TEMP. 130 °F
GAS VOL. 35 C.F.
H₂O 0.2 LBS
PARTIC 0 LBS
SO₂ 10 LBS
NONE
TEMP 130 °F
GAS VOL 35 CF
H₂O 0.2 LBS
PARTIC 0 LBS
SO₂ 10 LBS
EVAPORATORS
WATER = 2,223 LBS.
SOLIDS = 2,223 LBS.
WATER = 18,032 LBS
SO₂ = 5 LBS
TEMP 350
FLUE GAS VOL 122,100
LBS FLUE GAS 5,710
GRAINS/ACF 0.13
GRAINS/SDCF 0.26
PARTIC 2.28
SO₂ 1.10
MOISTURE 909
TEMP 350 °F
FLUE GAS VOL 122,100 CF
LBS FLUE GAS 5,710
GRAINS/ACF 0.66
GRAINS/SDCF 1.32
PARTIC 11.40 LBS
SO₂ 1.10 LBS
MOISTURE 909 LBS
D. CO
80%
BARK = 1,100 LBS
BAR
COMB
C 26.5 %
H 8.3 %
O 63.5 %
S 0.1 %
N 0 %
INERTS 1.6 %
BTU/LB 4,500
INCINER
COMBI
BO
C 45.4 %
H 5.5 %
O 34.3 %
S 9.1 %
N 4.6 %
INERTS 1.1 %
BTU/LB 8,500
(SEE NOTE 7)
LBS.
LBS.
LBS
LBS
WATER
SULFUR = 206 LBS.
MP 100 °F
S VOL. 15,300 CF
43 LBS
412 LBS
SO₂ GAS
SULFUR BURNER
AIR
LEGEND
PROCESS EQUIPMENT
CONTROL EQUIPMENT
AIR EMISSIONS
SUL
BATCH
PLANT AND L
EMIS
DE
CON
ENVIRONM

B


CHEMICALS, SOLIDS=100 LBS.
WATER = 1,800 LBS.
WATER =75,000 LBS
STEAM =1,000 LBS
BLEACH PLANT
B.D. PULP=1,760 LBS. @12%
WATER=12,900 LBS
SOLIDS=15 LBS.
WATER: 77,304 LBS
SOLIDS=144 LBS.
NONE
VAPORATORS
WATER = 2,223 LBS.
SOLIDS=2,223 LBS.
INCINERATE IN COMBINATION BOILER
BARK = 1,100 LBS
BARK
OIL/COAL
COMBINATION BOILER
D. COLL. 80% EFF.
OIL
COAL
NONE
D. COLL. 90% EFF.
D. COLL. 98% EFF
(SEE NOTE 7)

TEMP	350°F
FLUE GAS VOL	122,100CF
LBS FLUE GAS	5,710
GRAINS/ACF	0.13
GRAINS/SDCF	0.26
PARTIC	2.28LBS
SO_2	1.10LBS
MOISTURE	909LBS

	No 6 OIL	COAL
TEMP	350°F	350°F
FLUE GAS VOL	228,500CF	227,500CF
LBS FLUE GAS	11,200	11,410
GRAINS/ACF	0.02	0.19
GRAINS/SDCF	0.04	0.32
PARTIC	0.68	6.23
SO_2	27.20	35.60
MOISTURE	775	576

TEMP	350°F
FLUE GAS VOL	427,000CF
LBS FLUE GAS	20,000
GRAINS/ACF	0.05
GRAINS/SDCF	0.12
PARTIC	3.20 LBS
SO_2	400 LBS
SO_3	UNKNOWN LBS
MOISTURE	3,520 LBS

TEMP	350°F
FLUE GAS VOL	122,100CF
LBS FLUE GAS	5,710
GRAINS/ACF	0.66
GRAINS/SDCF	1.32
PARTIC	11.40LBS
SO_2	1.10LBS
MOISTURE	909LBS

	No 6 OIL	COAL
TEMP	350°F	350°F
FLUE GAS VOL	228,500CF	227,500CF
LBS FLUE GAS	11,200	11,410
GRAINS/ACF	0.02	1.91
GRAINS/SDCF	0.04	3.19
PARTIC	0.68	62.30
SO_2	27.20	35.60
MOISTURE	775	576

C	26.5%
H	8.3%
O	63.5%
S	0.1%
N	0%
INERTS	1.6%
BTU/LB	4,500

	#6 OIL	COAL
LBS	680	890
C	87.4%	72.0%
H	10.5%	5.5%
O	.0%	10.5%
S	2.0%	2.0%
INERTS	0.1%	10.0%
BTU/LB	18,300	14,000

TEMP	350°F
FLUE GAS VOL	427,000CF
LBS FLUE GAS	20,000
GRAINS/ACF	0.44
GRAINS/SDCF	0.91
PARTIC	26.70 LBS
SO_2	400 LBS
SO_3	UNKNOWN LBS
MOISTURE	3,520 LBS

TEMP	130°F
GAS VOL	35 CF
H_2O	0.2 LBS
PARTIC	0 LBS
SO_2	10 LBS

C	45.4%
H	5.5%
O	34.3%
S	9.1%
N	4.6%
INERTS	1.1%
BTU/LB	8,500

NOTES

1. ALL FLOWS REPRESENT THOSE FOR ONE AIR DRY TON OF UNBLEACHED PULP.
2. THE EMISSIONS FROM BURNING BARK AND OIL/COAL ARE BASED ON SUPPLYING THE TOTAL STEAM-ELECTRICAL REQUIREMENTS FOR PULP AND PAPER PRODUCTION (SEE "POWER PLANT ENERGY BALANCE"). OF THESE EMISSIONS, APPROXIMATELY 28% CAN BE ATTRIBUTED TO PAPER PRODUCTION AND 72% TO PULP PRODUCTION.
3. THE BARK FLOW IS BASED ON DEBARKING 100% OF THE WOOD SUPPLY. FOR MILLS RECEIVING WOOD CHIPS, THE BARK FLOW WILL BE REDUCED IN PROPORTION TO THE AMOUNT OF CHIPS BEING RECEIVED. THEREFORE, TO SATISFY THE STEAM REQUIREMENTS ON THE "POWER PLANT ENERGY BALANCE", THE OIL AND COAL FLOWS MUST BE INCREASED ACCORDINGLY.
4. ABBREVIATIONS:
    - C. F. AND A. C. F. ACTUAL CUBIC FEET AT FLOW CONDITIONS
    - S. D. C. F. = STANDARD DRY CUBIC FEET AT 70° F. AND 14.7 P.S.I.A.
    - EFF. = ANNUAL OPERATING EFFICIENCY AT THE CONTROL EQUIPMENT
5. ALL VALUES ARE ANNUAL AVERAGES.
6. ALL FUEL ANALYSES ARE ON AN "AS FIRED" BASIS.
7. CLEMENTS, J. L. AND SAGE, W. L., "AMMONIUM LIQUOR BURNING AND SO_2 RECOVERY", PRESENTED AT 23rd ENGINEERING CONFERENCE OF TAPPI, NOVEMBER , 1968.

LEGEND
▭ PROCESS EQUIPMENT
▢ CONTROL EQUIPMENT
◯ AIR EMISSIONS

SULFITE FLOW DIAGRAM No. 4

BATCH DIGESTER, AMMONIUM ACID SULFITE, WITH BLEACH PLANT AND LIQUOR INCINERATION WITHOUT CHEMICAL RECOVERY

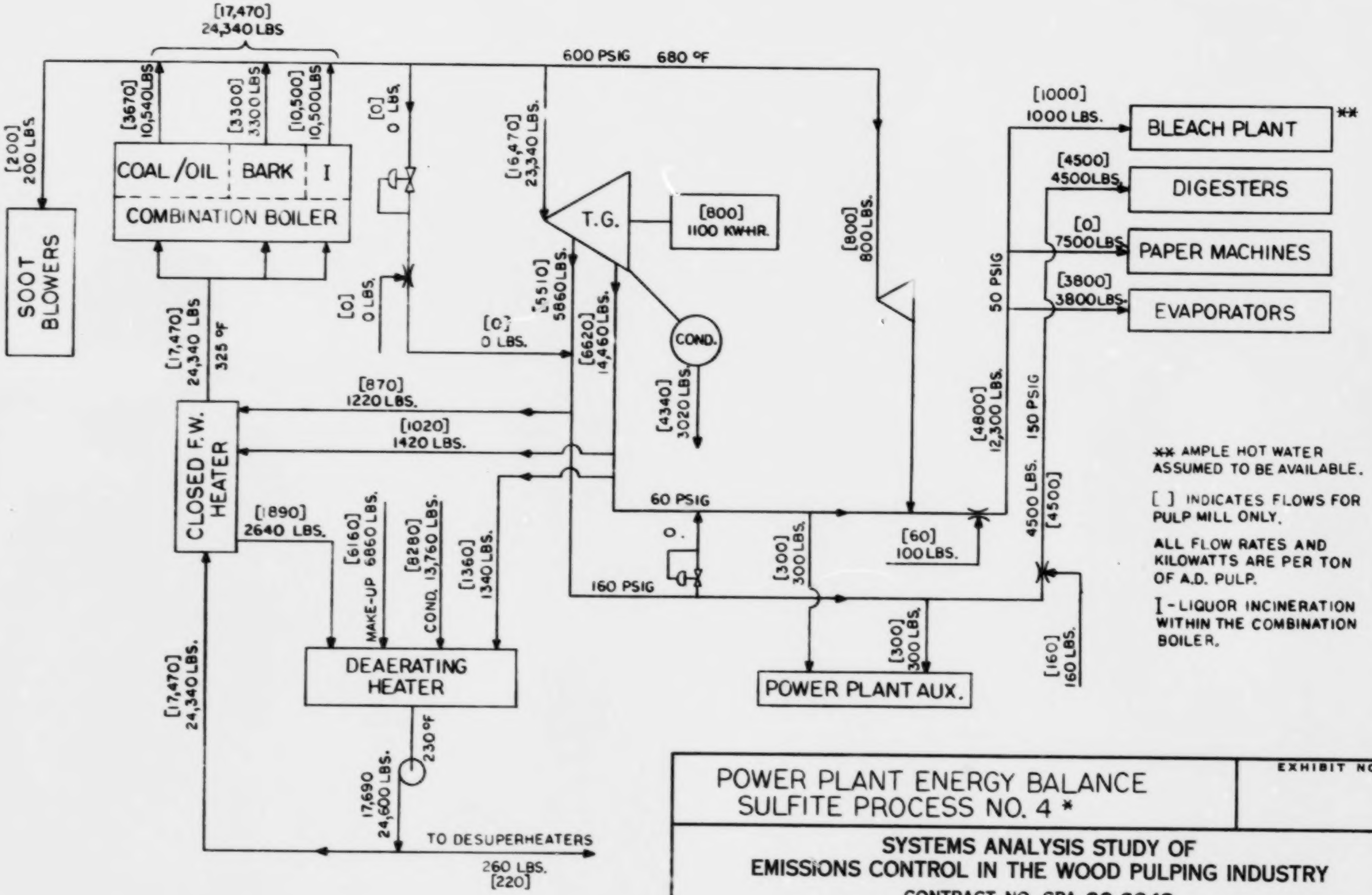
EXHIBIT NO.

SYSTEMS ANALYSIS STUDY OF EMISSIONS CONTROL IN THE WOOD PULPING INDUSTRY

CONTRACT NO. CPA 22-69-18
FOR
DEPARTMENT OF HEALTH, EDUCATION AND WELFARE
CONSUMER PROTECTION AND ENVIRONMENTAL HEALTH SERVICE
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

ENVIRONMENTAL ENGINEERING, INC.
GAINESVILLE, FLORIDA

J. E. SIRRINE COMPANY, *ENGINEERS*
GREENVILLE, S. C.


[17,470]
24,340 LBS
600 PSIG 680 °F
[200]
200 LBS
SOOT BLOWERS
[3670]
10,540 LBS
[3300]
3300 LBS
[10,500]
10,500 LBS
COAL/OIL
BARK
I
COMBINATION BOILER
[0]
0 LBS.
[0]
0 LBS.
[0]
0 LBS.
[16,470]
23,340 LBS.
T.G.
[800]
1100 KW-HR.
[5510]
5860 LBS.
[6620]
14,460 LBS.
COND.
[4340]
3020 LBS.
[800]
800 LBS.
[17,470]
24,340 LBS
325 °F
[870]
1220 LBS.
[1020]
1420 LBS.
CLOSED F.W. HEATER
[1890]
2640 LBS.
[6160]
MAKE-UP 6860 LBS.
[8280]
COND. 13,760 LBS.
[1360]
1340 LBS.
60 PSIG
0.
160 PSIG
[300]
300 LBS.
[300]
300 LBS.
POWER PLANT AUX.
[60]
100 LBS.
[4800]
12,300 LBS.
50 PSIG
150 PSIG
4500 LBS.
[4500]
[160]
160 LBS.
[1000]
1000 LBS.
BLEACH PLANT **
[4500]
4500 LBS.
DIGESTERS
[0]
7500 LBS.
PAPER MACHINES
[3800]
3800 LBS.
EVAPORATORS
DEAERATING HEATER
230 °F
[17,470]
24,340 LBS.
17,690
24,600 LBS.
TO DESUPERHEATERS
260 LBS.
[220]

** AMPLE HOT WATER ASSUMED TO BE AVAILABLE.

[ ] INDICATES FLOWS FOR PULP MILL ONLY.

ALL FLOW RATES AND KILOWATTS ARE PER TON OF A.D. PULP.

I - LIQUOR INCINERATION WITHIN THE COMBINATION BOILER.

* BATCH DIGESTERS, AMMONIUM ACID SUFITE, WITHOUT CHEMICAL RECOVERY, BLEACHING, AMMONIUM LIQUOR INCINERATION BLEACH PLANT.

POWER PLANT ENERGY BALANCE
SULFITE PROCESS NO. 4 *

EXHIBIT NO.

SYSTEMS ANALYSIS STUDY OF
EMISSIONS CONTROL IN THE WOOD PULPING INDUSTRY
CONTRACT NO. CPA 22-69-18
FOR
DEPARTMENT OF HEALTH, EDUCATION AND WELFARE
CONSUMER PROTECTION AND ENVIRONMENTAL HEALTH SERVICE
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ENVIRONMENTAL ENGINEERING, INC.
GAINESVILLE, FLORIDA

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GREENVILLE, S. C.

CHAPTER IV

QUANTITY AND NATURE OF EMISSIONS

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CHAPTER 4

QUANTITY AND NATURE OF EMISSIONS

SUMMARY

The control of gaseous and particulate emissions from the various processes in chemical wood pulping requires an understanding of the quantity and nature of the compounds involved. This information is limited. The largest amount of data is available for the kraft process. Because of the numerous variables which affect emissions, it is virtually impossible to give more than a broad range of values without monitoring specific sources in individual mills.

For each pulping process considered in this study, both gaseous and particulate emissions are discussed. The nature and effects of the compounds of interest are discussed briefly. Then for each source, where the information is available, the theory of production is discussed, the operating and process variables which affect emissions are described, an assessment of the relative importance of the process as a source is made, ranges of emissions for major compounds are cited, and an attempt is made to estimate the lowest emission which can be attained by optimum operation. There are numerous gaps in this format simply because the necessary information was not available.

4.1 INTRODUCTION

Planning for the control of gaseous and particulate emissions from various processes requires an understanding of the quantity and nature of the compounds involved. Although the wood pulping industry has been engaged in research efforts with respect to its air quality improvement programs for many years, valid information on emissions is rather sparse. The condition is due in part to slow development of dependable analytical techniques. In many instances, the low concentrations of gases are near the limit of resolution of most standard analytical techniques. The largest amount of information on emissions is available for the various modifications of the kraft process.

4.1.1 SOURCES OF DATA

Information on the chemical nature and formation of particulate emissions from various pulping operations has been known for some time and is readily available in the literature. Only in recent years, however, have definite studies been made to explain the production and identify more than broad categories of gaseous emissions. Although the recent literature contains some data of this nature, much information had to be obtained by direct communication.

Quantitative emission data were obtained from the literature, from in-house studies, and by direct communication.

4.1.2 LIMITATIONS OF DATA

Attempting to set forth specific quantitative values for emissions from a process as complicated as chemical wood pulping involves considerable risk. Process and operating variables may have profound and sometimes unknown effects on emissions. Wide and relatively rapid fluctuations may occur in some operating variables and in emissions. A number of combinations of unit processes is possible in a mill to produce the same end result. The nature and manner of operation of each unit process has an effect on the emissions from every other unit process. In addition, unit processes frequently are modified from their original design.

Only the main compounds of interest have been monitored although numerous others may be present. Reliable analytical techniques are not available for all compounds of interest. Even in those instances where so called reasonably reliable sampling and analytical techniques have been in use for some time, the actual procedure selected will influence the apparent concentration.

For these reasons it is virtually impossible to give more than a broad range of values without monitoring specific sources in specific mills.

4.1.3 BENEFITS TO BE OBTAINED BY MONITORING SOURCES

Only by monitoring individual sources in specific mills can reasonably reliable values be obtained for the range of concentration for compounds of interest. Procedures for such monitoring and the limitations of the procedures are discussed in Chapter 9. Variations in emissions need to be correlated with operating variables.

It must be realized that reliable sampling and analytical procedures are not available for all sources and all compounds which may be of interest. Despite the limitations of existing analytical procedures, monitoring of individual sources as an operational guide may result not only in reduced emissions but in increased efficiency of operation.

4.2 GASEOUS EMISSIONS FROM THE KRAFT PROCESS

4.2.1 GASEOUS EMISSIONS FROM THE KRAFT PROCESS - GENERAL

The characteristic kraft mill odor is principally due to the presence of a variable mixture of hydrogen sulfide, methyl mercaptan, dimethyl sulfide, and dimethyl disulfide. All of these gases contain sulfur, which is a necessary component of the kraft cooking liquor. Hydrogen sulfide, methyl mercaptan, dimethyl sulfide, and dimethyl disulfide are referred to as reduced sulfur compounds, and the latter three gases are usually described as organosulfur compounds.

Hydrogen sulfide emissions are derived from breakdown of the weak base, sodium sulfide, which is the characteristic of kraft cooking liquor. It may also be generated by improper operation of a recovery furnace. Methyl mercaptan and dimethyl sulfide are formed in reactions with the wood component lignin. Dimethyl disulfide is formed through the oxidation of mercaptan groups derived from the thiolignins.

A great deal of variability exists in the published values of the odor thresholds for these sulfur gases, as can be seen in Table 4-1. The detection of these gases by the olfactory senses is obviously a highly individualistic determination.

Table 4.2 presents some of the physical properties of these sulfur gases.

Sulfur Dioxide (SO_2). Sulfur dioxide emissions in the kraft process result from oxidation of reduced sulfur compounds. A potential source of sulfur dioxide is the recovery boilers, where reduced sulfur gases present can be oxidized in the furnace atmosphere.

Sulfur dioxide is a mildly acidic gas and is readily absorbed by the alkaline black liquor in the direct contact evaporator. The mechanism is an acid-base reaction:

$$SO_2 + 2NaOH \rightleftarrows Na_2SO_3 + H_2O \qquad \text{Eq. 4.1a}$$

$$Na_2SO_3 + SO_2 + H_2O \rightleftarrows 2NaHSO_3 \qquad \text{Eq. 4.1b}$$

This reaction will, of course, result in a lowering of the pH of the liquor, because of the reduction in free alkali.

TABLE 4-1

ODOR THRESHOLDS OF KRAFT MILL GASEOUS SULFUR COMPOUNDS IN AIR

ppm (by volume)

Sulfur Dioxide	1.0-5.0	(2)				
Hydrogen Sulfide	0.0085	(1),	0.0047	(2),	0.0009	(4)
Methyl Mercaptan	0.0021	(2),	0.040	(3),	0.0006	(4)
Dimethyl Sulfide	0.0001	(2),	0.0036	(3),	0.0003	(4)

TABLE 4-2

CHARACTERISTICS OF KRAFT MILL GASEOUS SULFUR COMPOUNDS

Compound	Characteristic Odor	Color	Explosive Limits (air) Lower	Upper	
Sulfur Dioxide	strong, suffocating	none	Not-explosive		
Hydrogen Sulfide	rotten eggs	none	4.3%	46%	(5)
Methyl Mercaptan	rotten cabbage	none	3.9%	22%	(6)
Dimethyl Sulfide	vegetable sulfide	none	2.2%	9%	(6)

The principal potential sources of sulfur dioxide in the kraft mill are the power boilers and the recovery furnace. Emissions from the power boilers depend upon the type of fuel and the sulfur content of the fuel. With a 2 percent sulfur fuel oil, the sulfur dioxide emissions range from 12 - 43 lb/ADT of pulp; for a 2 percent sulfur coal, the emissions range from 15 - 57 lb/ADT of pulp. The sulfur dioxide emissions from the recovery furnace range from 0.5 - 15 lb/ADT of pulp.

Hydrogen Sulfide (H_2S). Hydrogen sulfide is a feebly acidic gas which partially ionizes in aqueous solution. The ionization proceeds in two stages with the formation of hydrosulfide and, with increasing pH, sulfide ions

$$H_2S \rightleftarrows HS^- + H^+ \rightleftarrows S^= + 2H^+ \qquad \text{Eq. 4.2}$$

$$\text{increasing pH} \rightarrow$$

Black liquor contains a high concentration of dissolved sodium sulfide in strongly alkaline solution. If the pH were depressed, the sodium sulfide would hydrolyze to sodium hyrosulfide and below pH 8 appreciable unionized hydrogen sulfide would form as the reaction equilibria in Eq. 4.2 moves from right to left. Shih (28) reports that at a pH of about 8.0, most hydrogen sulfide forms hydrosulfide ions, so in normal black liquor conditions, there is very little dissolved hydrogen sulfide in the liquor.

Due to the equilibrium between the hydrosulfide ion and water vapor, hydrogen sulfide gas can be stripped from black liquor at steam vents. There could be, therefore, a problem in the evaporator areas of the kraft mill.

Hydrogen sulfide is formed in the recovery furnace in the reducing atmosphere found in the lower sections, as the sulfur containing compounds from the black liquor are volatilized and reduced. However, in normal furnace operation the hydrogen sulfide oxidizes to sulfur dioxide (which is largely absorbed within the furnace) in the combustion sections before the exhaust gases leave the furnace.

Hydrogen sulfide generally represents the largest gaseous emission from the kraft process. Two of the most effective means for reducing the hydrogen sulfide emissions from kraft mills are black liquor oxidation and maintaining tight control of critical process operating variables.

Methyl Mercaptan (MeSH). Methyl mercaptan is a reduced sulfur compound which is formed during the kraft cook by the reaction of hydrosulfide ion and the methoxy-lignin component of the wood (8):

$$\text{Lignin-OCH}_3 + \text{HS}^- \rightarrow \text{MeSH} + \text{Lignin-O}^- \qquad \text{Eq. 4.3}$$

Methyl mercaptan will also dissociate in an aqueous solution to methyl mercaptide ion. Shah (9) has reported that this dissociation is essentially completed above a pH of 12.0:

$$\text{MeSH} + \text{OH}^- \rightleftarrows \text{MeS}^- + \text{HOH} \qquad \text{Eq. 4.4}$$

Methyl mercaptan will, therefore, be present in low concentrations as a dissolved gas in the black liquor. As the pH decreases, the equilibrium shown above shifts to the left and MeSH gas is evolved.

Methyl mercaptan is primarily emitted from the digester relief and blow where it is formed, and from the brown stock washers where the pH of the liquor drops below the equilibrium point. Its emission decreases as its residual concentration in the liquor diminishes.

Dimethyl Sulfide and Dimethyl Disulfide (MeSMe, MeSSMe). Dimethyl sulfide is primarily formed through the reaction of methyl mercaptide ion with the methoxy-lignin component of the wood (8). It does not, however, dissociate as hydrogen sulfide and methyl mercaptan do:

$$\text{Lignin-OCH}_3 + \text{MeS}^- \rightarrow \text{Lignin-O}^- + \text{MeSMe} \qquad \text{Eq. 4.5}$$

Dimethyl sulfide may also be formed by the disproportionation of methyl mercaptan. At normal liquor temperatures (150 - 200°F) it is highly volatile.

Dimethyl disulfide is formed by the oxidation of methyl mercaptan throughout the recovery system, especially in oxidation towers. Dimethyl disulfide has a higher boiling point than any of the other compounds and its retention in the liquor is therefore greater than the other organosulfur compounds.

$$4\text{MeSH} + \text{O}_2 \rightleftarrows 2\text{MeSSMe} + 2\text{H}_2\text{O} \qquad \text{Eq. 4.6}$$

4.2.2 GASEOUS EMISSIONS FROM THE KRAFT RECOVERY FURNACE

4.2.2.1 Formation of Gaseous Pollutants

The purposes of burning concentrated black liquor in the kraft recovery furnace are: the recovery of sodium and sulfur, the production of steam, and the disposal of unwanted dissolved wood components in the liquor. In most instances, liquor of 62% solids content or greater will burn in a self-supporting combustion. Sodium and sulfur can be recovered in the form of sodium sulfide and sodium carbonate.

The recovery furnace theoretically is divided into three sections: the drying and pyrolysis zone, the reducing zone, and the oxidizing zone. The black liquor is intorduced to the furnace through spray guns located in the drying zone. The heat in the furnace is sufficient to immediately evaporate the remaining water from the liquor and to cause the organic solids within the liquor to undergo pyrolysis. Pyrolysis is defined as the chemical change brought about by the action of heat upon a substance.

Pyrolysis occurs in the drying zone of the furnace. The drying zone of most recovery furnaces is in a combined oxidation and reduction state because of the manner in which the primary air is introduced into the furnace. Forced draft fans force the primary air into the furnace through the air ports located around the perimeter of the furnace at the level of the drying zone.

It is felt that sulfur which exists in organic substances undergoing pyrolysis in the oxidizing atmosphere will usually be oxidized to form nonvolatile sulfur radicals such as sulfite, thiosulfate, and sulfate (10). These non-volatile sulfur radicals in the form of sodium salts will fall into the reducing zone of the furnace, although small particles may be carried out of the furnace by the draft of the fan.

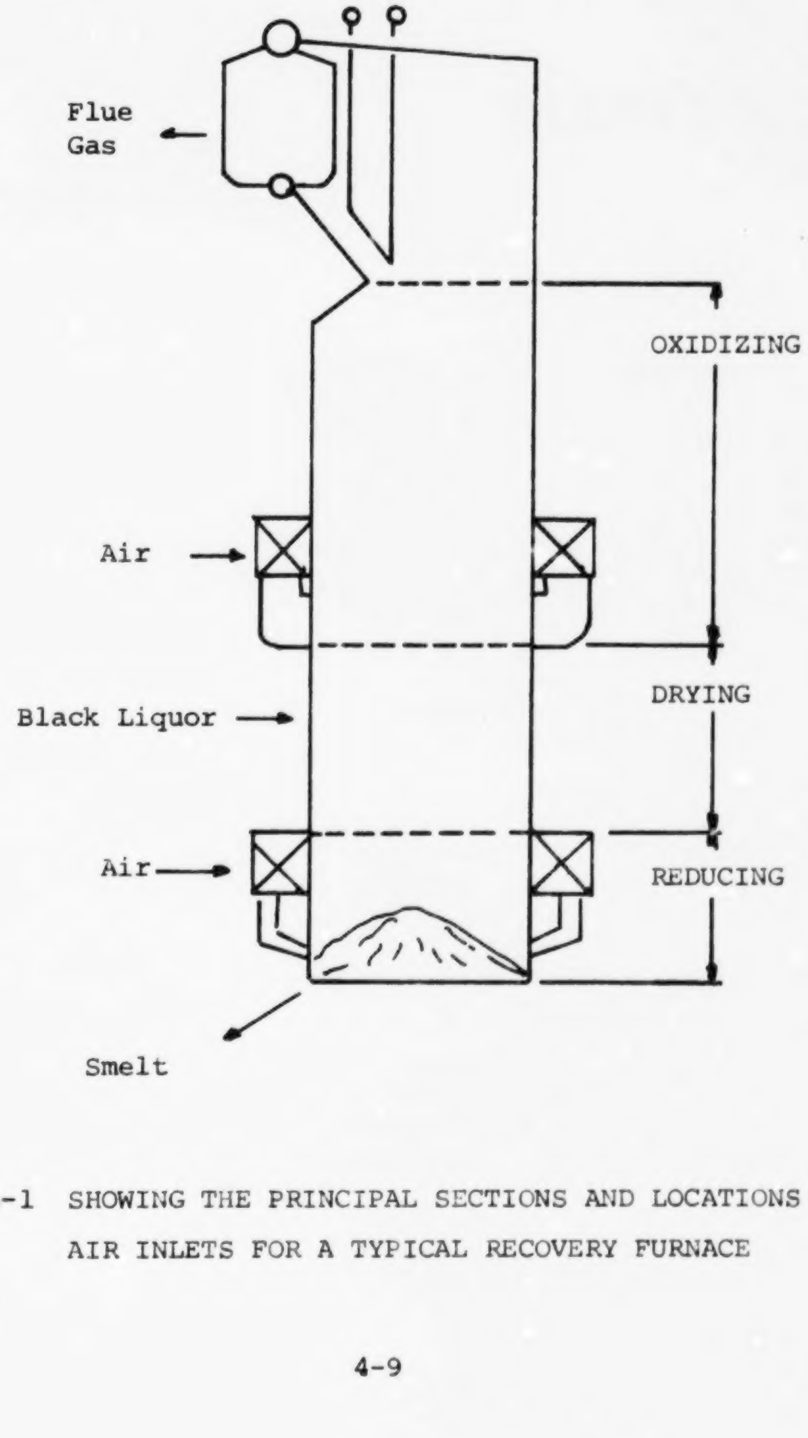

Flue
Gas
OXIDIZING
Air
DRYING
Black Liquor
Air
REDUCING
Smelt

FIGURE 4-1 SHOWING THE PRINCIPAL SECTIONS AND LOCATIONS OF AIR INLETS FOR A TYPICAL RECOVERY FURNACE

Conversely, sulfur which exists in organic substances undergoing pryolysis in a reducing atmosphere, may form volatile reduced sulfur compounds (10). These gases flow into the upper regions of the furnace.

These proposals exist only as theory, and much work remains to be done in understanding the mechanism of pyrolysis of black liquor before these theories can be regarded as correct.

. The sodium salts of the oxidized sulfur compounds which have fallen into the reducing zone along with the carbon ash residue will undergo reduction to sodium sulfide and sodium carbonate. Carbon dioxide and carbon monoxide are gaseous by-products of these reduction processes. The fused sodium sulfide and sodium carbonate are withdrawn and mixed with water in the smelt tank to form green liquor.

The total reduced sulfur (TRS) gases and the carbon monoxide which are drawn into the oxidizing zone of the furnace should undergo oxidation to sulfur dioxide, carbon dioxide, and water This oxidation will be carried to completion if proper conditions of temperature, excess oxygen, residence time and turbulence are provided. When these conditions do not exist, complete oxidation will not occur and the reduced sulfur compounds will escape from the furnace.

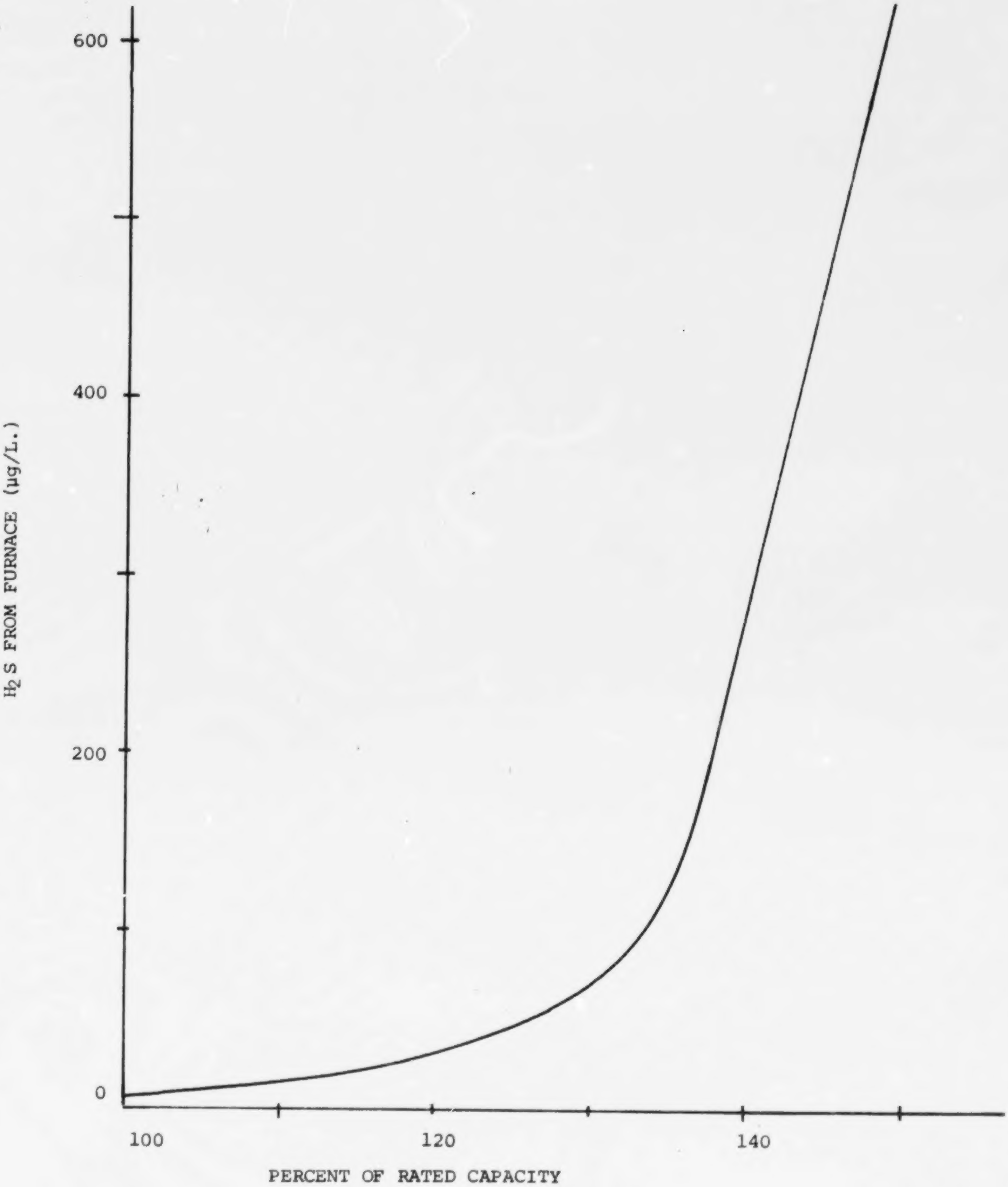
4.2.2.2 Effect of Operating Variables upon Emissions from the Recovery Furnace

There are many operating variables which have been shown to affect the emissions from a kraft recovery furnace. Several of those which are considered to be the most important and which are understood the best are: the rate at which black liquor is fired into the furnace, the ratio of secondary air to the black liquor firing rate, the percent excess oxygen in the furnace flue gas, black liquor spray droplet size, and turbulence within the recovery furnace.

Several authors have identified the relationship which exists between the firing rate of a furnace and the gaseous emissions (10, 11, 12, 13). Figure 4-2 shows the relationship between the firing rate and the hydrogen sulfide emissions from a particular recovery furnace. The hydrogen sulfide emissions gradually increase until a sharp upward break occurs. The point at which this sharp increase occurs is primarily dependant upon the capacity of the forced draft fan.

FIGURE 4-2

THE EFFECT OF OVERLOADING A KRAFT RECOVERY FURNACE UPON HYDROGEN SULFIDE EMISSIONS



The other operating variables discussed in this section are important in determining the rate at which the emissions will increase with overloading.

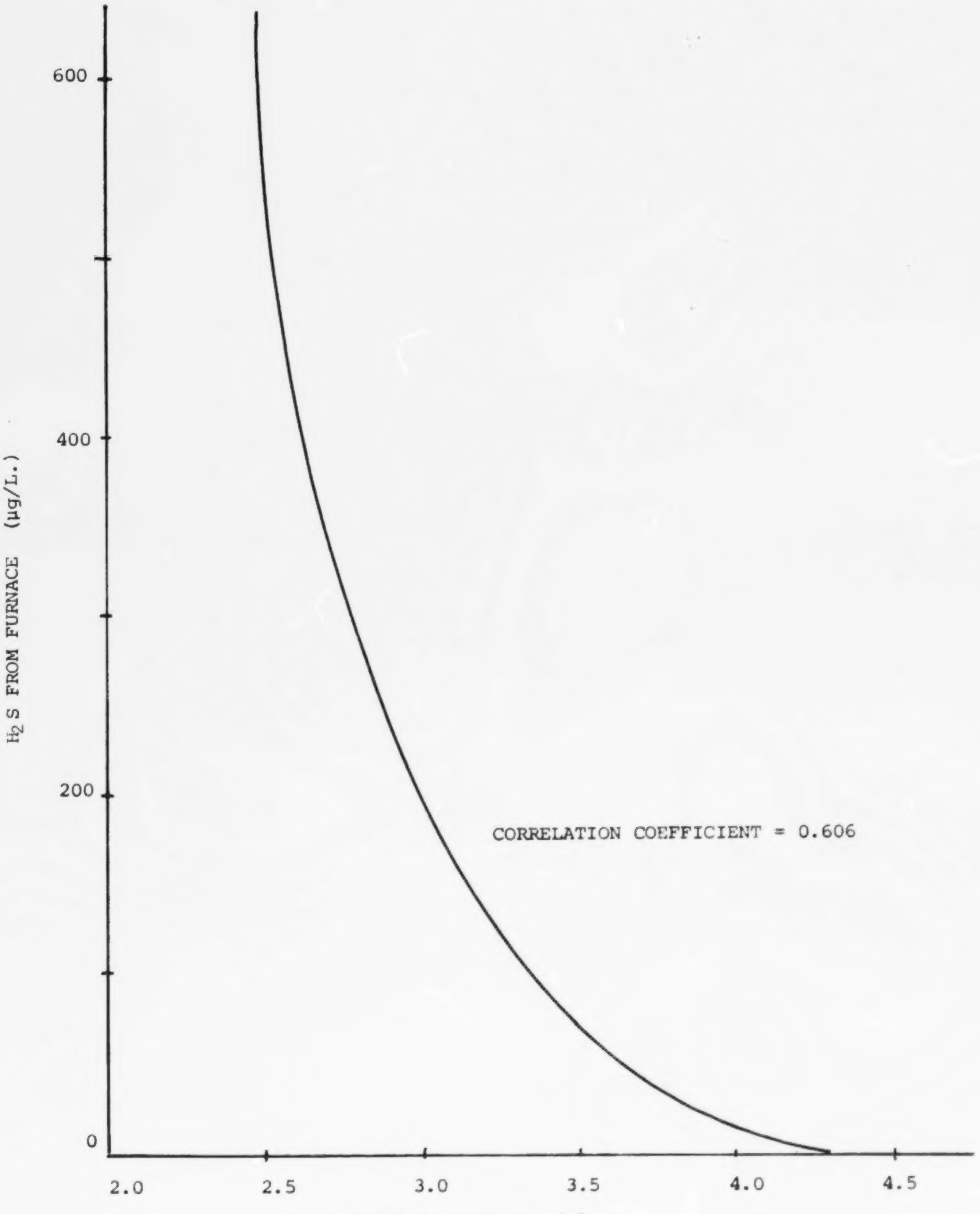
The I.D. fans installed on a furnace are usually conservatively designed and can handle small increases in the black liquor firing rate which are required during peak loading. However, a point will be reached when the fan is at its limiting capacity and it will no longer be able to provide sufficient oxygen to oxidize the rising gases (10). In addition, secondary air will not be introduced at a rate sufficient to provide adequate mixing to thoroughly oxidize the gases. Increasing the capacity of the fan will not solve the problem totally. As the air flow within the furnace is increased, carry-over of black liquor droplets will occur. These particles will burn in the upper sections of the furnace, generating excessive heat which may result in superheater tube failure.

The increased hydrogen sulfide emission of an overloaded furnace is therefore partially caused by the poor gas flow characteristics and lack of oxygen within the oxidizing zone of the furnace, both of which are required to maintain minimum emissions.

It has also been shown that to maintain minimum reduced sulfur emissions from a recovery furnace the percent of excess oxygen in the recovery flue gas must be maintained above a minimum level (10). The presence of sufficient excess oxygen does not, however, insure minimal emissions unless certain conditions exist in the furnace: the efficient mixing of the available oxygen with the combustible material, a temperature high enough to provide rapid chemical reaction, and sufficient residence time to allow oxidation to occur. Murray and Rayner (10) have experimentally determined in the laboratory the relationship between the percent excess oxygen in the flue gas and the hydrogen sulfide emissions as shown in Figure 4-3. Thoen (13) has also derived similar relationships, but the percent excess oxygen required for a particular furnace must be experimentally determined for each case. Most authors have reported that minimum TRS emissions occur when oxygen in the flue gas from the recovery furnace is about 2.5 - 4.0 percent by volume.

FIGURE 4 - 3

THE EFFECT OF RESIDUAL OXYGEN IN THE FLUE GAS UPON HYDROGEN SULFIDE EMISSIONS



The ratio of the secondary air flow rate to the black liquor firing rate is also an important variable. At the ratio which yields the lowest emissions, the excess oxygen requirement is usually satisfied (10, 13). Murray and Rayner's relationship of this ratio versus the hydrogen sulfide emissions is shown in Figure 4-4. Results by other authors show that the secondary air should amount to about 30 - 40 percent of the air supplied to the furnace.

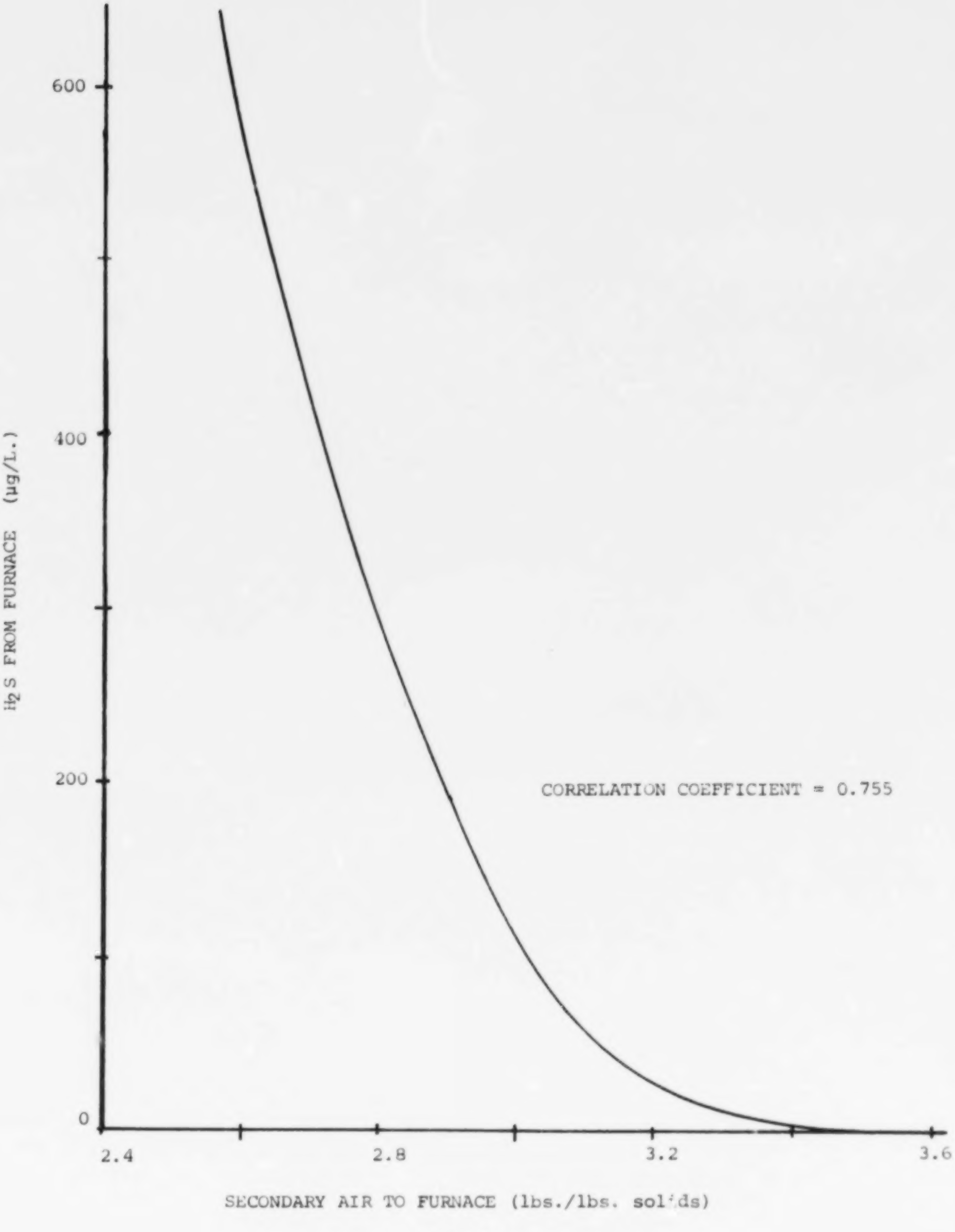
Thoen (13) has reported that when black liquor is sprayed into the furnace in small drops there exists a tendency for the solids to be carried up to the oxidizing zone immediately, whereas larger spray droplets will fall to the bottom of the furnace. When black liquor is carried to the upper zones it will burn slowly, yielding excessive amounts of heat which will overheat the boiler. The products of combustion will then be swept out of the furnace before they can be oxidized and emissions of TRS gases will be increased. Thoen has given TRS emission values (Table 4-3) for "coarse" and "fine" sprays which show about a 1000 percent increase between the two. Unfortunately the sizes of the coarse and fine spray droplets were not reported, but the nozzle sizes, line pressures and temperatures corresponding to the coarse and fine spray conditions are shown in Table 4-3. The size of the spray droplets can be controlled by the temperature and pressure of spraying the liquor and by the type of spray nozzles.

The turbulence created by the introduction of the secondary air is of great importance. Turbulence creates thorough mixing and thus more efficient oxidation of the gases. A well designed introduction system for the secondary air will provide a high degree of turbulence while imparting no vertical velocity to decrease residence time. The more common methods of injection are horizontally and tangentially arranged inlet ports.

The channeling of the reducing atmosphere into the upper regions of the furnace has two effects. First it allows an increased period of contact between sulfur containing organic solids and the reducing atmosphere which will lead to further reduced sulfur emissions. Secondly, it will decrease the efficiency of the oxidizing zone of the furnace. Channeling can be easily controlled by proper introduction of secondary air to create turbulence to mix the oxygen with the reducing atmosphere.

FIGURE 4 - 4

THE EFFECT OF THE SECONDARY AIR/BLACK LIQUOR FIRING RATE RATIO UPON HYDROGEN SULFIDE EMISSIONS



To date it has not been shown experimentally that there is a relationship between the sulfide ion concentration in the black liquor and the quantity of gaseous emission of any compound from the recovery furnace itself. Thus any decrease in emissions from the furnace itself while burning oxidized black liquor as opposed to unoxidized liquor is purely coincindental. The effect of oxidized and unoxidized liquor on reduced sulfur emissions shows up in the direct contact evaporator following following the recovery furnace, where emissions will be decreased if the liquor is oxidized.

The wide variety of recovery furnace designs makes it impossible to present data which can be considered to be typical for every furnace. The data and figures presented in this section are given to show trends which have been observed on selected furnaces.

TABLE 4-3

THE EFFECT OF SPRAY SIZE ON SULFUR GAS EMISSIONS FROM A KRAFT RECOVERY FURANCE*(13) (USING UNOXIDIZED BLACK LIQUOR)

	Flue Gas Concentration (ppm)				
	SO_2	H_2S	MeSH	MeSMe	MeSSMe
Coarse Spray	2.12	0.02	0	0	0
Fine Spray	10.70	2.40	0.12	0	0.03

*These data are representative of a particular furnace only.

Note: Coarse Spray - No. 4 Nozzle at 17 psi and 240°F

Fine Spray - No. 2 Nozzle at 17 psi and 245°F

4.2.2.3 Relative Importance of the Recovery Furnace

The importance of the kraft recovery furnace as a source of emissions is dependent upon the operation of the furnace itself. As has been described in this section, there exists an optimum set of operating conditions for each furnace which will result in the reduction of the emissions of reduced sulfur compounds to a negligible level. If the furnace is operated at conditions other than optimum, the quantity of emissions will increase as the divergence from the optimum condition increases. The kraft recovery furnace may then become the major source of sulfurous emissions in the mill.

4.2.2.4 Ranges of Emissions of Gaseous Sulfur Compounds

The importance of the operating variables has been determined. In general, recovery furnaces are rarely operated under optimum conditions. The ranges of emissions in Table 4-4 typify the emission ranges for kraft recovery furnaces operating under non-optimum conditions. Variances from these ranges may occur if the furnace is operated under severe operating conditions. Operation under optimum conditions is discussed in the following section.

TABLE 4-4

APPROXIMATE RANGES OF EMISSIONS FROM KRAFT RECOVERY FURNACES
(Before the Direct Contact Evaporator)

	lb/ADT of pulp
Sulfur Dioxide	10.0 - 15.0
Hydrogen Sulfide	1.0 - 5.0
Methyl Mercaptan	0.01- 0.10
Dimethyl Sulfide	0.01- 0.02
Dimethyl Disulfide	0.01- 0.02

4.2.2.5 Emissions Under Optimum Conditions

A kraft recovery furnace which is operated at optimum operating conditions which have been individually determined for the furnace in question can be expected to emit negligible amounts of gaseous sulfur compounds.

In his study of the effects of operating conditions upon emissions from the furnace, Theon (13), operated a recovery furnace under optimum conditions for a 24 hour period, and obtained the results in Table 4-5.

TABLE 4-5

KRAFT RECOVERY FURNACE
EXTENDED OPERATION AT OPTIMUM CONDITIONS* (13)
(24hr)

(load-116% design)

Periodic Samples	ppm v/v					%		
	SO_2	H_2S	RSH	RSR	RSSR	O_2	CO_2	CO
1	0.04	0	0	0	0	3.1	15.6	0
2	0.07	0	0	0	0	4.4	15.4	0
3	0.04	0	0	0	0	4.4	15.4	0
4	0.01	0	0	0	0	4.8	15.4	0
5	0.08	0	0	0	0	2.1	16.4	0

*35% secondary air at 180 ft/sec; coarse black liquor spray.

Note: (1) 0 indicates concentration less than detection limits of analytical equipment.

(2) Work currently under way indicates the sulfur dioxide concentrations presented in this table may be low due to the limitations of the analytical techniques used.

4.2.3 GASEOUS EMISSIONS FROM THE DIRECT CONTACT EVAPORATOR

4.2.3.1 Formation of Gaseous Pollutants

Gaseous emissions from the direct contact evaporator are caused by the stripping of the dissolved gases from the black liquor by the furnace flue gases. This occurs when a concentration difference exists between the actual concentration of the gases in the flue gas and the equilibrium concentration which is consistent with the temperature and pH of the liquor (14). In black liquor, the dissolved hydrogen sulfide and methyl mercaptan are normally low, as discussed in Section 4.2.1. However, the absorption of carbon dioxide and sulfur dioxide from the flue gas reduces the pH of the liquor causing an increase in the concentrations of these dissolved gases. This point will be elaborated upon shortly. Dimethyl sulfide and dimethyl disulfide usually have low residual concentrations in the unoxidized liquor and their emission from this source is characteristically low. The concentrations of these two gases in oxidized liquor is usually slightly higher (especially dimethyl disulfide) and the emissions may be slightly greater than for unoxidized liquor.

The absorption of carbon dioxide and sulfur dioxide takes place according to the straightforward acid-base equilibrium reactions:

$$OH^- + CO_2 \overset{\text{high pH}}{\rightleftarrows} H^+ + CO_3^= \qquad \text{Eq. 4.7}$$

$$OH^- + SO_2 \overset{\text{high pH}}{\rightleftarrows} H^+ + SO_3^= \qquad \text{Eq. 4.8}$$

In the case of black liquor, the carbonate and sulfite ions are both stronger acids than hydrosulfide and mercaptide ions and will displace the latter ions from the liquor. The acid hydrogen ions formed by the absorption of these gases will enter into competitive reactions between hydrosulfide, mercaptide and hydroxide ions in the liquor:

$$H^+ + HS^- \rightleftarrows H_2S\uparrow \qquad \text{Eq. 4.9}$$

$$H^+ + RS^- \rightleftarrows RSH\uparrow \qquad \text{Eq. 4.10}$$

$$H^+ + OH^- \rightleftarrows H_2O \qquad \text{Eq. 4.11}$$

In actuality, all three reactions take place. The proportion of the ion which enters into each reaction will depend upon the relative strengths of the negative ions listed and the concentrations of the ions.

Most of the hydrogen ion will probably react with the hydroxide ion to lower the pH of the solution. This shift in equilibrium will cause an increase in the dissolved concentration of hydrogen sulfide and methyl mercaptan gases in the liquor, and the reactions which take place in Eq. 4.9 and 4.10 will tend to increase emissions from the evaporator. The escaping vapor may also serve as a vehicle for hydrogen sulfide and methyl mercaptan stripping.

4.2.3.2 Effect of Operating Variables Upon Emissions from the Direct Contact Evaporator

For a given concentration of a dissolved gas in the black liquor, there exists an equilibrium concentration of that compound in the gas above the liquor, which is consistent with the temperature and the pressure of the system. According to the principles of mass transer, a transfer of mass from the liquor to the gas phase will occur if the concentration of the gas is lower than the equilibrium concentration.

Conversely, mass transfer will occur from the gas to the liquor if the concentration in this gas phase is greater than the equilibrium concentration.

Murray and Rayner (13) have developed a mathematical model which approximates the actual conditions for hydrogen sulfide transfer. Their formula shows that the equilibrium concentration of hydrogen sulfide above the liquor is dependent upon the sulfide ion concentration and the pH of the liquor. The mathematical model for the mass transfer of hydrogen sulfide is dependent upon this equilibrium concentration and the concentration of hydrogen sulfide in the entering gas stream as well as the gas flow rate. Tables 4-6 and 4-7 show the effect of pH and sulfide concentration upon the emission of hydrogen sulfide from the direct contact evaporator. The data are taken from the pilot plant work under controlled conditions.

A similar mathematical model could be developed for methyl mercaptan. Dimethyl sulfide and dimethyl disulfide do not dissociate in black liquor and their removal from the liquor is a function of their vapor pressures and the temperature of the liquor, rather than a function of the pH.

TABLE 4-6

THE EFFECT OF CHANGING THE pH OF THE BLACK LIQUOR ON HYDROGEN SULFIDE EMISSIONS DURING DIRECT CONTACT EVAPORATION (14)

(Pilot Plant Study)

pH	Na_2S Cone in Black Liquor (g/l.)	H_2S Cone in Recovery Furnace Flue Gases (μg/l.)	H_2S Cone in Contact Evaporator Exit Gas (μg/l.)	Change in H_2S Across Contact Evaporator (μ/l.)
12.59	14.2	16	53	37
12.33	18.3	41	185	144
12.07	15.3	36	273	237

TABLE 4-7

THE EFFECT OF CHANGING THE SODIUM SULFIDE CONCENTRATION BY OXIDATION ON THE EMISSION OF HYDROGEN SULFIDE DURING CONTACT EVAPORATION (14)

(Pilot Plant Study)

pH	Na_2S Cone in Black Liquor (g/l.)	H_2S Cone in Recovery Furnace Flue Gases (μg/l.)	H_2S Cone in Contace Evaporator Exit Gas (μg/l.)	Change in H_2S Across Contact Evaporator (μ/l.)
11.85	28.4	204	580	376
	20.2	Zero	216	216
	Zero	74	50	- 24
12.30	33.3	32	295	263
	18.3	41	186	145
	Zero	14	10	- 4

It should be noted that if black liquor oxidation is performed to a residual sulfide concentration of less than 0.1 gm/liter, the pH of the liquor is no longer a factor in emissions from the direct contact evaporator.

4.2.3.3 Relative Importance of the Direct Contact Evaporator

The significance of the direct contact evaporator as an emissions source is dependent upon the residual sodium sulfide concentration and the pH of the black liquor being handled. Hydrogen sulfide emissions have been observed to increase rapidly with increasing sodium sulfide residual concentrations in black liquor. High sodium sulfide concentrations in black liquor may result in the direct contact evaporator becoming a major source of emissions. When emissions are high, the hydrogen sulfide emissions may be decreased by maintaining a higher pH.

Complete oxidation (99+ percent) of kraft black liquor to produce a negligible sodium sulfide concentration will virtually eliminate the direct contact evaporator as a major emission source within the kraft mill.

4.2.3.4 Ranges of Emissions of Gaseous Sulfur Compounds

Because of the importance of the sodium sulfide concentration upon the emission of hydrogen sulfide and methyl mercaptan, ranges of emissions for both oxidized and unoxidized liquor have been tabulated. The ranges of emissions in Table 4-8 represent the sum of the recovery furnace and direct contact evaporator contributions, and do not separate the increase in emissions across the direct contact evaporator from the portion that may enter with the recovery furnace gases.

4.2.3.5 Emissions under Optimum Conditions

From the very low reduced sulfur emissions from a modern recovery furnace operated under optimum conditions, it is concluded that the emissions from a direct contact evaporator are primarily dependent upon the degree of black liquor oxidation. At 99+ percent oxidation efficiency, the emissions should be of the be of the order of 0.5 lb/ADT.

If high residual sulfide concentrations remain in the liquor during evaporation, the emissions can be controlled to a limited extent by maintaining the pH as high as is feasible.

TABLE 4-8

APPROXIMATE RANGES OF EMISSIONS FROM THE DIRECT CONTACT EVAPORATOR

	UNOXIDIZED LIQUOR (lb/ADT)	OXIDIZED LIQUOR (lb/ADT)
Sulfur Dioxide	2.0 - 8.0	2.0 - 8.0
Hydrogen Sulfide	5.0 - 30.0	0.10 - 2.0
Methyl Mercaptan	0.50 - 2.50	0.05 - 0.25
Dimethyl Sulfide	0.10 - 0.30	0.01 - 0.10
Dimethyl Disulfide	0.10 - 0.40	0.01 - 0.20

4.2.4 GASEOUS EMISSIONS FROM THE DIGESTER RELIEF AND BLOW

4.2.4.1 Formation of Gaseous Pollutants

In a digester, chips are treated with predetermined quantities of alkali, in the form of caustic sulfide liquors, and subjected to heat and pressure to separate fibrous constituents of wood by dissolving the nonfibrous constituents (15). Digester "relief is an essential part of [batch] digester operation; it is done for four purposes: circulation, control of cooking, reduction of digester pressure before blowing and removal of air. . .(16)." In some cases the gases which are relieved from the digester may be vented to a turpentine recovery system where the crude turpentine is condensed. Noncondensibles in the relief gases will pass through the turpentine condenser and are a possible source of emissions at this point unless further processing of them is arranged. When a turpentine recovery system is not used, the relief gases are sent to a heat recovery system, after which the noncondensible gases represent a potential emission problem. When the cook is completed, the contents of the digester are blown into the blow tank by the pressure within the digester. "The contents are blown tangentially into the top of the blow tank, the stock dropping into the tank and the steam and gases escaping from the top vent (17)." The steam and gases pass through a heat accumulator from where they may be emitted to the atmosphere.

The hydrolytic equilibria for the sulfide ions in the kraft liquor are:

$$S^{=} + HOH \overset{K_1}{\rightleftarrows} HS^{-} + OH^{-} \qquad \text{Eq. 4.12}$$

$$HS^{-} + HOH \overset{K_2}{\rightleftarrows} H_2S + OH^{-} \qquad \text{Eq. 4.13}$$

While previous work indicates that in kraft pulping, the equilibrium for Equation 4.13 lies almost completely to the left, reliable values for the equilibrium constant for Equation 4.12 are not available. However, K_1 is known to be large, and it is proper to express the total sulfur in the liquor as hydrosulfide ion (HS^{-}) (8).

The demethylation of lignin is believed to be accomplished by a nucleophlic attack by the hydrosulfide ions upon the methoxyl group of the lignin. In a consecutive bimolecular reaction the mercaptide ion attacks another methoxyl group and yields dimethyl sulfide (8).

$$\text{Lignin-OMe} + HS^- \rightarrow MeSH + \text{Lignin-O}^- \qquad \text{Eq. 4.14}$$

$$MeSH + OH^- \rightarrow MeS^- + HOH \qquad \text{Eq. 4.15}$$

$$\text{Lignin-OMe} + MeS^- \rightarrow MeSMe + \text{Lignin-O}^- \qquad \text{Eq. 4.16}$$

Thus since the methyl mercaptan concentration increases as a consequence of the reaction in Eq. 4.14, the rate of reaction 4.16 likewise increases, which tends to deplete mercaptan by conversion to dimethyl sulfide.

Ultimately, a steady state condition may be reached at which the mercaptan is formed and depleted at equal rates, and its concentration will remain constant. On the other hand the rate of dimethyl sulfide formation should be zero at the outset of the cook and gradually increase, approaching a steady rate of formation at steady state condition (see Figure 4-5) (8).

There are some side reactions which can occur during a cook. One is the oxidation of methyl mercaptan to dimethyl disulfide (8):

$$2RSH + 1/2\ O_2 \rightarrow RSSR + H_2O \qquad \text{Eq. 4.17}$$

It is conceivable that by eliminating oxygen from the system, this side reaction producing a malodorous gas could be suppressed.

Another side reaction which may be significant is the disproportionation of methyl mercaptan to dimethyl sulfide and hydrogen sulfide (8). This reaction has been found to have a slow but significant rate:

$$2RSH \rightarrow RSR + H_2S \qquad \text{Eq. 4.18}$$

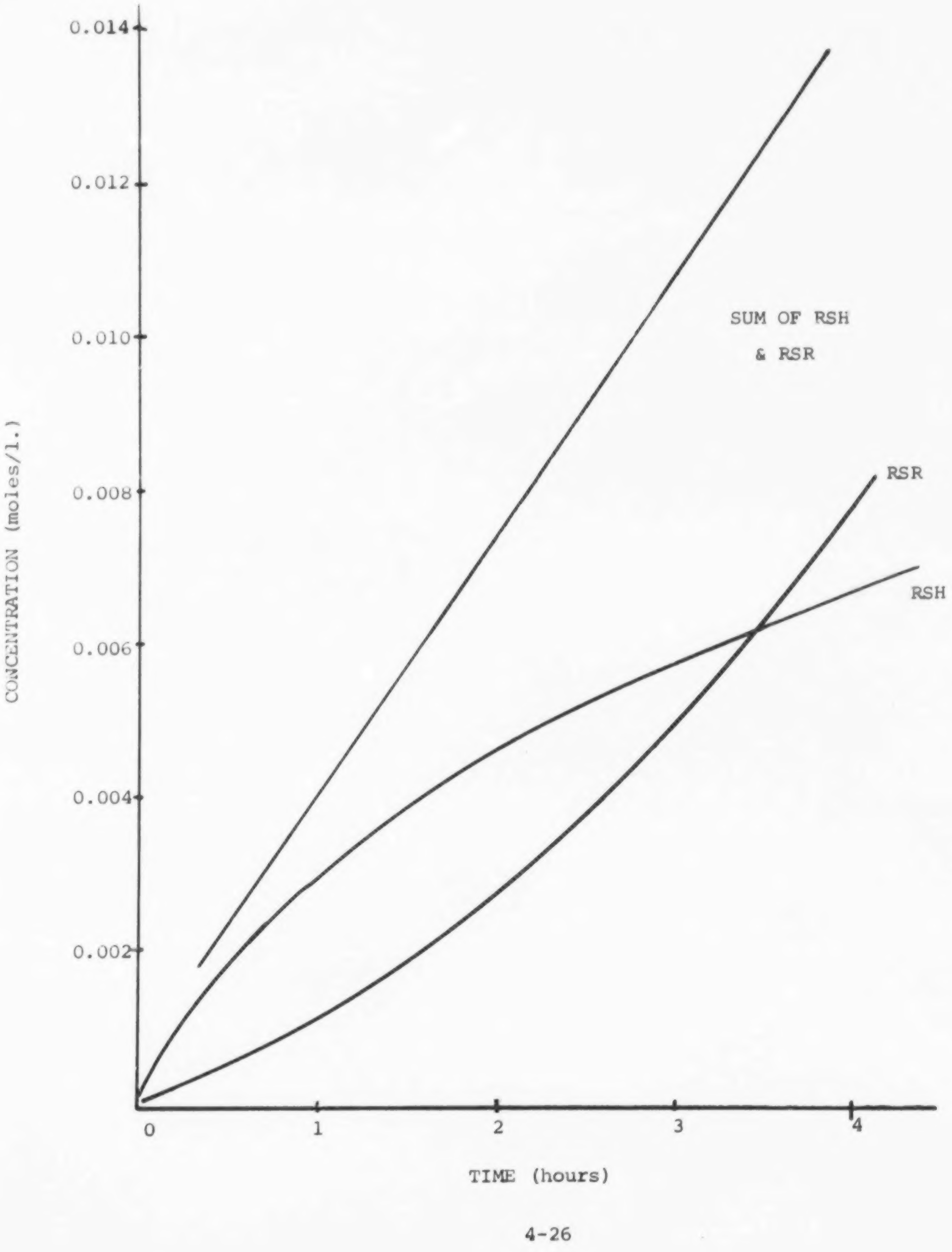
4.2.4.2 The Effect of Operating Variables Upon Emissions

McKean, Norwicki and Douglass have stated that minimum black liquor recycle, short duration cooks, low cooking liquor sulfidity and high residual alkali will help yield minimal emissions from the kraft digester. However these authors differ in their opinion of the effect of temperature in the digester upon emissions.

In the following paragraphs the relationships between various operating variables and emissions are discussed. It must be remembered, however, that practical verification has not been reported in many instances between the relationships established and variations in pulp quality produced under these conditions. In other words, it frequently is not possible to optimize operating conditions to produce minimum emissions and still produce pulp of the desired quality.

FIGURE 4 - 5

THE RATE OF FORMATION OF METHYL MERCAPTAN AND
DIMETHYL SULFIDE DURING DIGESTION AT 180° C



Effect of Wood Species on Odor Formation. Investigations into the quantitative amounts of gaseous products from the digesters have indicated that the amounts of TRS produced from hardwood cooks are greater than those produced from softwood cooks under the same conditions (7, 8). In the case of hardwoods, there is an initial rapid rate of production of malodors that ultimately approaches a constant rate that is about 10 percent greater than softwoods. As a result, a 4-hour hardwood cook will produce about 30 percent greater quantity of odorous compounds than a softwood cook (7, 8). The production of a greater quantity of odorous compounds indicates that the syringyl methoxyls from hardwoods react 10 percent faster than quaiacyl methoxyls which are found in the softwoods (7, 8).

Rate of Mercaptide Ion Attack on Lignin Methoxyl. Two hypotheses have been formed to account for the slow initial rate of dimethyl sulfide formation. The first is that as the pulping proceeds the transfer of lignin into the soluble phase from the solid phase would enhance the accessibility of the lignin to mercaptide ion attack. However, this hypothesis has been partially disproved by showing that carbohydrate-free lignin preparations, which would have better accessibility, still suffer from slow initial attack by mercaptide ions (8).

The second hypothesis is accounted for by purely chemical factors. The structural changes which take place in the lignin to cause solubilization during pulping, cause severe degradation to lower weight fragments. It may, therefore, be expected that these changes would effect the methoxyl reactivity (8).

Temperature Dependance of Reactions. Experimental laboratory investigation has derived the activation energies for the principle reactions occuring during pulping (7). The activation energies for the odor forming reactions (Eq. 4.14 and 4.16) are 7.6 and 11.3 kcal/mole, while the activation energy of the delignification has been estimated to be about 30 kcal/mole.

By inspection of the Arrhenius equation, which describes the temperature dependence of the rate constant, it appears that an increase in temperature will accelerate the delignification reaction while having only a moderate effect upon the odor forming reactions.

Thus, for example, if the cooking temperature were raised from 160 to 190°C the rate of delignification would be increased by a factor of 10 while the odor forming reaction rate would be increased by a factor of about 2.3. Thus for the same amount of delignification the odorous compounds formed during the 190°C cook would be only one-fourth of the amount produced at 160°C.

It is worth noting that at elevated temperatures the formation of odorous gases (see Eqs. 4.14 and 4.16) takes place at a greater rate. The practice of blowing a digester at the earliest time which is consistant with the pulp quality desired is practiced for economic reasons. However, the importance of practicing this proceedure from an emissions standpoint can be seen in the previous example.

A continuous digester can eliminate these problems because the temperature of the pulp can be raised and lowered quickly, and the dissolved lignins can be removed before they enter the activated forms which eventually cause odors.

The Effect of Sulfidity. Because of the ionic equilibrium between sulfide ions and hydrosulfide ions, a higher liquor sulfidity will result in higher hydrosulfide ion concentrations. The net result will be a greater formation of methyl mercaptan by the reaction shown in Equation 4.14.

The higher hydrosulfide ion concentrations will also result in greater hydrogen sulfide emissions created by steam stripping, although this problem appears to be of less importance than the additional methyl mercaptan formation mentioned previously. In general, higher sulfidities will result in greater losses of reduced sulfur compounds.

The Effect of pH. The pH of the cooking liquor has an effect upon the emission of hydrogen sulfide and methyl mercaptan, as discussed in Section 4.2.1. Because methyl mercaptan is readily evolved at a pH below about 12, it is desirable to maintain a high residual pH after the cook to minimize this volatilization. A high pH, however, will not prevent the stripping of these gases (7). The feasibility of controlling the pH in this instance is controversial.

Heat Accumulator Size. In cases where direct contact heat accumulators are used, the volume of water through which the digester blow gases pass in the heat accumulator has a great effect upon the emissions of two of the reduced sulfur gases. Hydrogen sulfide and methyl mercaptan have limited solubilities in water (0.035 lb/gal and 0.187 lb/gal respectively at 70°F), and the greater the volume of water through which the noncondensible gases pass, the greater the absorption of them will be.

4.2.4.3 Relative Importance of the Digester Relief and Blow Emissions

The digester relief and blow may be the largest source of organosulfur emissions within the kraft mill. This is because these compounds are primarily formed in the digesters and the first opportunity for stripping from the liquor is at this point. At process points beyond this the residual organosulfur compounds will be stripped to a lesser extent.

The digester is a relatively minor source of hydrogen sulfide emissions when compared with the recovery furnace, direct contact evaporators, multiple effect evaporators, and the lime kilns. Sulfur dioxide emissions are almost negligible from the digester blow and relief.

4.2.4.4 Ranges of Emissions from the Digester Relief and Blow

Taking into account the variability of the operating conditions, the ranges of emissions for gaseous sulfur compounds from batch digesters are listed in Table 4-9. Emission data from continuous digesters are not currently available and the emissions from such equipment must not be considered identical to batch digester emissions.

TABLE 4-9

APPROXIMATE RANGES OF EMISSIONS FROM THE DIGESTER RELIEF AND BLOW

	lb/ADT
Sulfur Dioxide	Trace - 0.01
Hydrogen Sulfide	0.01 - 0.12
Methyl Mercaptan	0.02 - 0.40
Dimethyl Sulfide	0.40 - 2.5
Dimethyl Disulfide	0.20 - 1.50

4.2.5 GASEOUS EMISSIONS FROM THE LIME KILN SYSTEM

4.2.5.1 Formation of Gaseous Pollutants

There is little published information concerning gaseous emissions from the lime kiln. It has been considered to be a minor source of sulfur gases and thus few studies have been made.

Possible sources of sulfur compounds into the kiln system include the fuel used to fire the unit, residual concentrations of reduced sulfur compounds in the lime mud, noncondensible gases burned in the kiln, and scrubbing liquor used in the kiln scrubbers.

Taylor theorizes (19) that a part of the gaseous reduced sulfur emissions from the kiln will depend upon the residual concentration of the reduced sulfur compounds and sodium sulfide concentration in the lime mud. Thus the thoroughness of washing the mud to remove these residual concentrations could greatly affect the emissions. Unfortunately no quantitative data exists on the residual concentration and emission relationship. The mechanisms by which reduced sulfur compounds may be produced and their subsequent conversion are unknown.

Combustion of the noncondensible reduced sulfur compounds resulting from other processes in the lime kiln has been used with considerable success. In the hot end of the kiln, these gases as well as those mentioned in the previous paragraph are subjected to temperatures in the range of 1500-1800°F. Under these conditions, with sufficient excess oxygen, oxidation of the reduced sulfur gases takes place with a high degree of efficiency. The critical temperature and oxygen ranges are unknown.

Sulfur dioxide which would be a gaseous product of such oxidation reactions as well as of the fuel oil combustion will be subject to immediate chemical absorption by the calcium carbonate or oxide. The scrubbers used on all kiln systems also are effective gas removal devices. Very little sulfur dioxde is emitted from the kiln system for this reason.

Taylor (19) has observed that where scrubbing liquor containing sulfides, such as weak wash, filtrates or evaporator condensates from unoxidized black liquor, is used in the lime kiln scrubbers hydrogen sulfide, methyl mercaptan, dimethyl sulfide and dimethyl disulfide may be stripped. He noted that evaporator condensates from well-oxidized weak black liquor do not result in any significant increase in odor level.

4.2.5.2 Effect of Operating Variables upon Emissions

Data on this are not available.

4.2.5.3 Relative Importance of Lime Kiln System Emissions

Data not available

4.2.5.4 Ranges of Emissions from the Lime Kiln System

Warther and Amberg (27) have reported a range of 0.01 - 0.83 lb/ADT of total reduced sulfur emissions from the lime kilns of four different pulp mills.

More complete data are not now available although work is currently under way to determine such information.

4.2.5.5 Emissions Under Optimum Conditions

Data not available.

4.2.6 GASEOUS EMISSIONS FROM THE MULTIPLE EFFECT EVAPORATOR

4.2.6.1 Formation of Gaseous Pollutants

Concentration of black liquor from 13 - 16 percent solids to 48 - 55 percent is almost always carried out with a multiple effect evaporation system. A multiple effect evaporator arrangement serves as an economical means for accomplishing this because in general one pound of steam will evaporate four to five pounds of water. Usually, five or six evaporation units (effects) make up the system. Each effect consists of a vapor head and a heating element. Hot vapors from the vapor head of a previous effect pass to the heating element of the following effect. The effects are operated at successively lower pressures which causes a decrease in the boiling point of the liquor. Vapors after the final effect are condensed in a condenser rapidly enough to maintain a high vacuum. A typical multiple effect evaporator is shown in Figure 4-6.

The emissions from the multiple effect evaporators are noncondensible reduced sulfur gases which are vaporized or stripped during the boiling. These noncondensible gases, with vapors created during boiling, pass to the heating element of the following effect. In order to eliminate an accumulation of noncondensible gases in the heating element each heating element is provided with a gas vent. The vents from the heating elements that are under a pressure greater than atmospheric are vented to its vapor head. The vents from the heating elements under a vacuum are usually valved to a common header going directly to either a barometric or surface condenser.

With a barometric condenser in use a limited quantity of hydrogen sulfide and methyl mercaptan gases are soluable in the water spray. The condensate will then become a potential water pollution problem. A small steam jet is used to remove noncondensible gases. If a surface condenser is used the noncondensible gases are separated from the condensibles. A small steam jet is used to remove the noncondensible gases. It is these noncondensible gases that create the problem of air pollution from multiple effect evaporators.

The reduced pressure in the latter effects will result in a higher evolution of the reduced sulfur compounds. This increased evolution and the steam stripping of the reduced

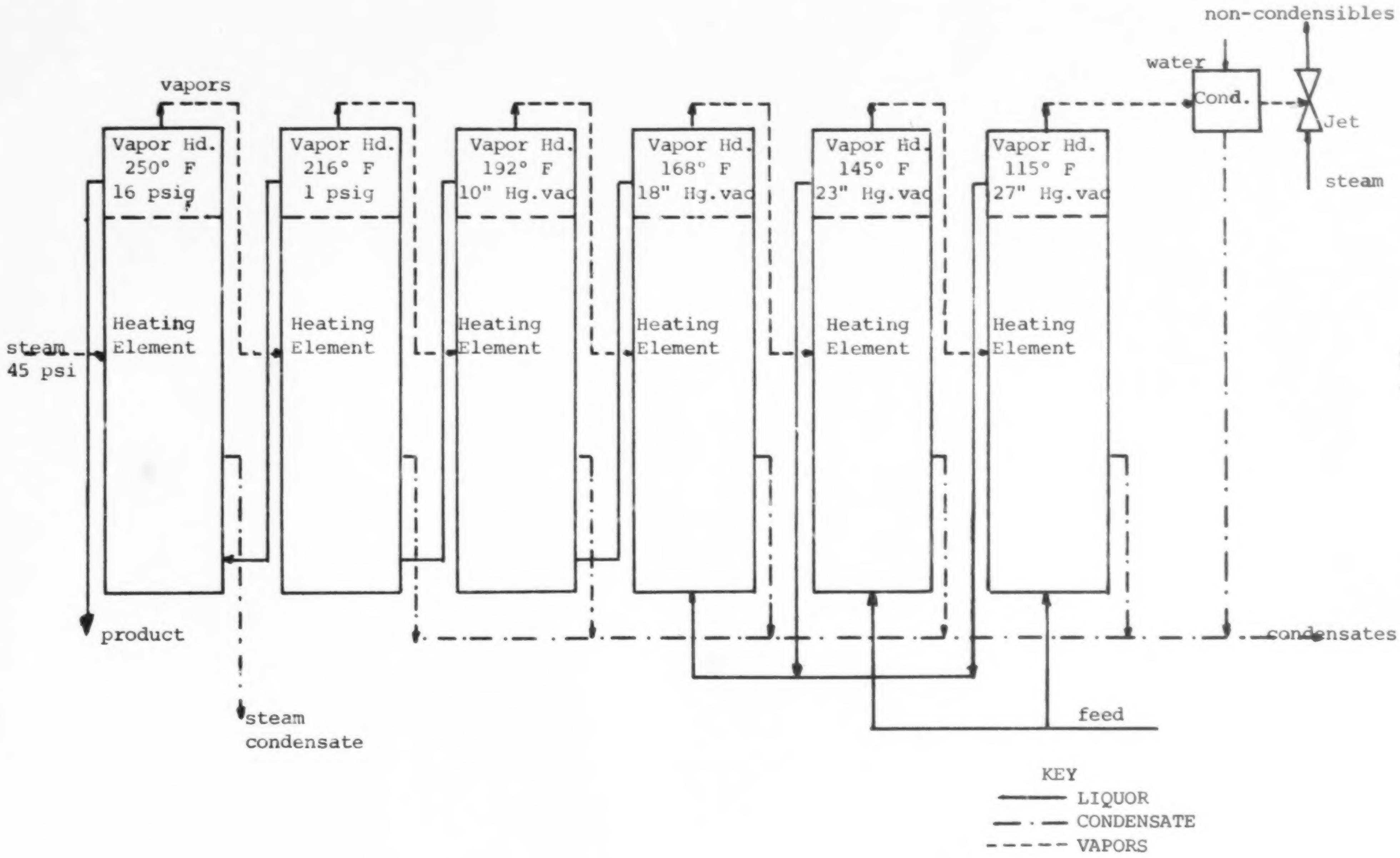

non-condensibles
water
Cond.
Jet
steam
vapors
Vapor Hd.
250° F
16 psig
Vapor Hd.
216° F
1 psig
Vapor Hd.
192° F
10" Hg.vac
Vapor Hd.
168° F
18" Hg.vac
Vapor Hd.
145° F
23" Hg.vac
Vapor Hd.
115° F
27" Hg.vac
Heating Element
Heating Element
Heating Element
Heating Element
Heating Element
Heating Element
steam
45 psi
product
condensates
steam
condensate
feed
KEY
LIQUOR
CONDENSATE
VAPORS

FIGURE 4-6 A TYPICAL MULTIPLE EFFECT EVAPORATOR SYSTEM

sulfur compounds are responsible for the emissions from the multiple effect evaporators. The steam stripping will be enhanced by the creation of foam within the evaporator tubes, because the foam will present a greater interfacial surface area for stripping.

4.2.6.2 Effect of Operating Variables Upon Emissions

Because emissions from the evaporators are caused by the steam stripping of reduced sulfur compounds and by the direct volatilization of these compounds, the sulfidity and the pH of the liquor will tend to be controlling factors in the quantity of gases emitted. The effects of these variables have been discussed in Section 4.2.1 and 4.2.3.

If weak black liquor oxidation has been performed, the emissions will be reduced because of the removal of the reduced sulfur compounds.

An important factor entering into the quantity of emissions from the evaporators when unoxidized black liquor is being processed, appears to be the type of condenser utilized. Harding (20) presented data on the emissions using two different types of condensers. These data are presented in Table 4-10.

Barometric condensers provide a reduction in the emission of reduced sulfur gases, but provide a contaminated condensate which may pose a water pollution problem. These reduced sulfur gases which have been scrubbed may be stripped from the condensates if the condensates are later used in the process.

Surface condensers provide a more efficient means for the collection and later destruction of the noncondensible gases.

TABLE 4-10

GASEOUS EMISSIONS FROM MULTIPLE EFFECT EVAPORATORS (20)

(lb/ADT)

Condenser Type	H_2S	RSH	RSR	RSSR
Surface	4.80	1.44	0.35	0.62
Barometric (with loss at noncondensible jet)	0.04	0.03	0.04	0
Barometric (losses from hot well)	0.13	0.20	0.14	0.02

4.2.6.3 Relative Importance of the Multiple Effect Evaporator

Noncondensible emissions from multiple effect evaporators are low volume, highly concentrated streams, and represent a major emission source within a kraft mill. Although black liquor oxidation may reduce the emissions significantly, the concentrations of reduced sulfur gases is such that further processing of the noncondensible gases is required. Overloading the evaporators should not increase the emissions from the evaporator per evaporated pound of solids introduced.

4.2.6.4 Ranges of Emissions from the Multiple Effect Evaporators

TABLE 4-11

APPROXIMATE RANGES OF EMISSION FROM THE MULTIPLE EFFECT EVAPORATORS

	Oxidized Black Liquor (lb/ADT)	Unoxidized Black Liquor (lb/ADT)
Sulfur Dioxide	0.01	0 - 0.01
Hydrogen Sulfide	0.01 - 0.02	0.10 - 3.0
Methyl Mercaptan	0.10 - 0.30	0.10 - 1.50
Dimethyl Sulfide	0.05 - 0.15	0.05 - 0.08
Dimethyl Disulfide	0.05 - 0.15	0.01 - 0.02

4.2.6.5 Emissions Under Optimum Conditions

Because of the wide variation in the design of multiple effect evaporators (i.e., the number of effects, type of evaporator use, feed locations, steam characteristics, and the variety of condensers utilized) the variability of conditions makes it impossible to define an optimum set of conditions for which to predict emissions.

4.2.7 GASEOUS EMISSIONS FROM THE BLACK LIQUOR OXIDATION TOWER

4.2.7.1 Formation of Gaseous Pollutants

Black liquor oxidation accomplishes the goal of converting the volatile reduced sulfur compounds of the black liquor to non-volatile or less volatile states. Thus the sulfide ions of the liquor are converted to thiosulfate ions, and the organosulfur compound are oxidized to dimethyl disulfide. Oxidation units come in a wide variety of styles, some of the more common styles are thin film towers, packed towers, and bubble trays. All oxidation units attempt to provide intimate contact between liquor and air while providing ease of handling black liquor.

Gaseous emissions from the oxidation tower are created by the stripping of the reduced sulfur compounds from the liquor by the air passing through it. Comparatively large volumes of dimethyl disulfide are emitted from this source, which may be explained by three factors. The first is the residual concentration of dimethyl disulfide in the liquor, the second is the fact that the methyl mercaptan and dimethyl sulfide are oxidized to dimethyl disulfide in the liquor, and the third is that methyl mercaptan and dimethyl sulfide which have been stripped into the gas may be oxidized in the gaseous states to dimethyl disulfide. Because dimethyl disulfide is the least volatile form of the organosulfur compounds present in the liquor, it would be preferable to accomplish the oxidation of the organosulfurs to dimethyl disulfide as quickly as possible to retain as much of the sulfur in the oxidation tower as possible.

4.2.7.2 Effect of Operating Variables Upon Emissions

The changes in operating variables which may favorably effect emissions from the oxidation tower may also have a detrimental effect upon the efficiency of the black liquor oxidation. However, there has not been enough work done in this area to discuss these effects thoroughly. It has been established that the temperature of the liquor must be held above about 160°F in order to prevent formation of elemental sulfur. Higher temperatures could result in a more efficient oxidation operation (power requirements, as well as overall efficiency), however, a higher temperature would mean greater volatility of the reduced sulfur compounds, and hence greater emissions. Another variable would be the air flow rate through (or over) the liquor. (From a given flow rate of air the effect of increasing the flow rate of air upon the oxidation efficiency and the emissions is not established.)

Black liquor oxidation is designed to decrease the emissions from the direct contact evaporator. For this reason the effects of the variables mentioned above upon the emissions from the oxidation tower must be balanced against the effect that a greater oxidation efficiency might have upon the emissions from sources effected by liquor oxidation. In other words, a slight increase in emissions from the oxidation tower created by conditions necessary for more efficient oxidation might result in a decrease in emissions throughout the mill of significantly greater magnitude. However, this balance must be studied in much greater detail on individual plants before any quantitative relationship can be developed.

In oxidation towers where oxidation is not completed to a high percentage (99+ percent), the residual reduced sulfur gases may be stripped from the liquor by the gases being passed through it. In a two stage oxidation system, the first stage would involve such a case.

4.2.7.3 Relative Importance of the Oxidation Tower

The organosulfur emissions from the oxidation tower, particularily the dimethyl disulfide emissions rank as the third largest source of emissions of this type behind the digester relief and blow, and the multiple effect evaporators. The emission of hydrogen sulfide and sulfur dioxide emissions from this source is negligible and the oxidation tower as a unit is considered a small source of atmospheric emissions.

4.2.7.4 Ranges of Emissions from the Oxidation Towers

TABLE 4-12

APPROXIMATE RANGES OF EMISSION FROM OXIDATION TOWERS

	(lb/ADT)
Sulfur Dioxide	0.00 - 0.01
Hydrogen Sulfide	0.01 - 0.02
Methyl Mercaptan	0.05 - 0.10
Dimethyl Sulfide	0.02 - 0.08
Dimethyl Disulfide	0.05 - 0.15

4.2.7.5 Emissions Under Optimum Conditions

This information is not available at the present time.

4.2.8 GASEOUS EMISSIONS FROM THE STOCK WASHERS

4.2.8.1 Formation of Gaseous Pollutants

Emissions from the brown stock washers arise primarily from the vaporization of the volatile reduced sulfur compounds. There are no chemical reactions which take place. However, there is a shift in the equilibrium for hydrogen sulfide and methyl mercaptan. Since the washer water has approximately a neutral pH, the dilution of the liquor by the water will cause a lowering of the pH to approximately 10.0. This pH is below the equilibrium point for methyl mercaptide ion and results in a corresponding shift to methyl mercaptan gas. The lower pH will also cause an increase in the concentration of dissolved hydrogen sulfide. However the equilibrium point of H_2S (8.0) will not be reached.

It is therefore reasonable to expect a large proportion of the emissions from the washers to be methyl mercaptan.

4.2.8.2 Effect of Operating Variables Upon Emissions

There are no well defined relationships for the effect of operating variables upon emissions from the washers, but several operations can be discussed quantitatively.

The amount of air drawn over the stock washers will affect the emissions because greater air flows will present greater driving forces for the mass transfer of the dissolved gases into the surrounding atmosphere.

The temperature of the water is also important because the volatility of the hydrogen sulfide and methyl mercaptans increases with temperature. The pH of the water also affects the emissions as previously discussed. The turbulence of the mixing of pulp and washer water may cause increased interface between the atmosphere and the liquid resulting in additional volatilization.

The thoroughness with which the liquor is removed in the first stage washer is important. If the liquor is efficiently removed in the first stage, the pH of the resulting weak liquor may remain sufficiently high to limit hydrogen sulfide emissions. There would also be less liquor in the remaining stages.

4.2.8.3 Relative Importance of the Emissions from the Stock Washers

The emissions from the stock washers are extremely low in all cases with the possible exception of methyl mercaptan.

4.2.8.4 Ranges of Emissions from the Stock Washers

TABLE 4-12

APPROXIMATE RANGES OF EMISSIONS FROM THE BROWN STOCK WASHERS

	lb/ADT
Sulfur Dioxide	0.01 - 0.02
Hydrogen Sulfide	0.01 - 0.12
Methyl Mercaptan	0.10 - 0.25
Dimethyl Sulfide	0.01 - 0.02
Dimethyl Disulfide	0.01 - 0.02

4.2.8.5 Emissions Under Optimum Conditions

The data are not available at present.

4.2.9 GASEOUS EMISSIONS FROM THE SMELT DISSOLVING TANK

4.2.9.1 Formation of Gaseous Pollutants

The gaseous emissions from the smelt dissolving tank are reduced sulfur compounds. Because the organosulfur compounds could not exist in the smelt at the smelt temperature, their presence in the vent gases must be accounted for by the introduction from outside sources. One such source might be the drafting of gases from the reducing zone of the furnace through the smelt spout of the furnace into the smelt tank. The natural draft of the smelt tank might cause a sufficient draft upon the furnace gases to pull these through a partially filled spout. Another source of organosulfur compounds might be the water used for smelt dissolving. In some instances, condensate water bearing these compounds is introduced to the causticizing system by way of the lime kiln scrubber and subsequently reach the smelt tank. A study of the water cycle from the scrubber will show that a portion of these condensates may eventually enter the smelt tank (see Kraft Flow Diagrams, Chapter Three) where the temperature would be favorable for volatization. However, the effect of such condensates in the smelt tank is believed to be very small.

Of course the chemical equilibrium of sulfide ion created by the dissociation of sodium sulfide, will create a condition for hydrogen sulfide emission resulting from stripping.

4.2.9.2 Effect of Operating Variables upon Emissions

Because the smelt tank is essentially an uncontrolled reaction vessel, there are no operating variables which can effect the emissions. However, proper design of the smelt spout could alleviate some of the emissions resulting from gases drawn from the recovery furnace.

4.2.9.3 Relative Importance of the Smelt Tank

The smelt tank is a minor source of emission of all gaseous compounds emitted from the kraft recovery system.

4.2.9.4 Ranges of Emissions from the Smelt Tank

TABLE 4-14

APPROXIMATE RANGES OF EMISSIONS FROM THE SMELT TANK

	(lb/ADT)
Sulfur Dioxide	0
Hydrogen Sulfide	0.02 - 0.05
Methyl Mercaptan	0.02 - 0.05
Dimethyl Sulfide	0.01 - 0.02
Dimethyl Disulfide	0.0 - 0.01

4.2.9.5 Emissions Under Optimum Conditions

The data are not currently available.

4.3 PARTICULATE EMISSIONS FROM THE KRAFT PROCESS

4.3.1 PARTICULATE EMISSIONS FROM THE KRAFT PROCESS - GENERAL

Particulate emissions from the kraft process occur primarily from the recovery furnace, the lime kiln, and the smelt dissolving tank. They are caused mainly by the carry-over of solids plus the sublimation and condensation of inorganic chemicals. The sublimation and condensation normally produces a fume. Little information is avaialble on the actual range of particle sizes from these sources, especially in the recovery furnace when agglomeration tends to occur readily. In addition, particulate emissions occur from combination and power boilers.

4.3.2 PARTICULATE EMISSIONS FROM THE RECOVERY FURNACE AND DIRECT CONTACT EVAPORATORS

4.3.2.1 Particulate Formation

Particulate emissions from the recovery furnace and direct contact evaporator complex consist primarily of sodium sulfate, and sodium carbonate. These emissions may result from both of the processes previously described. The high flue gas velocity may cause the carry-up of small droplets of black liquor which have been sprayed into the furnace. These droplets should burn in the oxidizing zone, but some of them may escape from the furnace.

The inorganic sodium salts which are found in particulate emissions from the furnace may be carried up by the furnace draft or may be formed in a vaporization - condensation process.

There are no particulates contributed by the direct contact evaporator. In fact, the evaporator may serve as a particulate reduction device, the method of reduction depending upon the type of evaporator in use.

4.3.2.2 Effect of Operating Variables Upon Emission

The portion of the sodium salt particles created by the sublimation-condensation process is not controllable because the initiating step, sublimation, takes place on the smelt pile where control is virtually impossible.

The portion of the particulate emissions which are attributable to the carry-up of black liquor solids is controllable to a small extent. The amount of solids carried up by the induced draft is dependent upon the spray droplet size and the vertical velocity in the furnace (see Section 4.2.2.2). Thus a large spray droplet size will reduce the carry-up of solids within the furnace.

Once in the oxidizing zone of the furnace, the solids will burn to gaseous and solid products. The solid products will initially consist of sodium sulfate, sodium carbonate, and organic solids. The sodium carbonate may react with the flue gas sulfur dioxide and oxygen to form additional sodium sulfate. The extent of this conversion will depend upon the sulfur dioxide concentration in the gas.

The organic solids will undergo a combustion-oxidation reaction. In most instances the oxidation will not be 100 percent efficient and some carbon residue will escape from the furnace. The operating conditions of temperature, percent excess oxygen, turbulence, and residence time will affect the efficiency of this reaction.

As mentioned in the previous section, the particulate concentration may be reduced when the flue gas passes through the direct contact evaporator. In a cascade evaporator a high flue gas velocity will provide a greater impingement rate for the particulates than a lower velocity. A Venturi evaporator can actually serve as a particulate scrubber.

Particulate emissions from the recovery furnace-direct contact evaporator complex are dealt with more effectively with particulate control devices such as electrostatic precipitators and Venturi scrubbers. It would be more advantageous to control furnace operating conditions to minimize gaseous emissions rather than particulate emissions. Although most conditions resulting in reduced gaseous emissions also serve to reduce particulate emissions.

4.3.2.3 Relative Importance

The kraft recovery furnace is the largest source of particulate emissions in the pulping system. Particulate control devices currently in use remove 80 - 99 percent plus of the particulates from the flue gas.

4.3.2.4 Ranges of Emissions

The range of particulate emissions from the recovery furance and direct contact evaporator (cascade and cyclone) is 75 - 125 lb/ADT without control devices. Using a 90 percent venturi evaporator scrubber after the furnace, the particulate emissions are in the range of 2- - 40 lb/ADT.

4.3.2.5 Emissions Under Optimum Conditions

These data are not currently available.

4.3.3 PARTICULATE EMISSIONS FROM THE LIME KILN

4.3.3.1 Particulate Formation

Particulate emissions from the lime kiln consist of the sodium salts, calcium carbonate, calcium sulfate, calcium oxide, and insoluble ash. The presence of the sodium salts may be accounted for by the sublimation-condensation process described in Section 4.3.2.1 and by dust entrainment within the kiln. However, neither calcium carbonate, nor calcium oxide will vaporize at temperatures within the kiln and the presence of calcium carbonate and calcium sulfate must be explained by the entrainment of the calcium carbonate and calcium oxide. The particles of calcium oxide may react with either the carbon dioxide or sulfur dioxide within the kiln to yield the appropriate calcium salt. Calcium carbonate may react with the sulfur dioxide and then oxygen to yield calcium sulfate.

$$CaO + CO_2 \quad CaCO_3 \qquad \text{Eq. 4.19}$$

$$CaO + SO_2 \quad CaSO_3 \qquad \text{Eq. 4.20}$$

$$CaCO_3 + SO_2 \quad CaSO_3 + CO_2 \qquad \text{Eq. 4.21}$$

$$CaSO_3 + 1/20 \quad CaSO_4 \qquad \text{Eq. 4.22}$$

The presence of the ash may stem from the burning of the fuel oil in the kiln.

4.3.3.2 Effect of Operating Variables upon Emissions

Other than controlling the rate of material output there appears to be no way of controlling the particulate emission for the lime kiln through manipulation of operating variables.

4.3.3.3 Relative Importance of Lime Kiln Particulate Emissions

The particulate emissions from the lime kiln (before the control equipment) are about one-fourth the emissions from recovery furnace but are still significantly large. However, the relatively low gas flow rates through the lime kiln allow installation of moderately priced high efficiency wet scrubbers which can reduce particulate emissions to low levels. At the present time, most lime kilns are equipped with wet scrubbers whose efficiencies range from 80 - 99 percent on the calcium salts. The particulate emissions from the lime kilns whose scrubber efficiencies are in the lower part of the range may be significant.

4.3.3.4 Ranges of Emission from the Lime Kiln

The range of particulate emission from the lime kiln before the particulate control equipmentis 20 - 65 lbs/ADT.

4.3.3.5 Emission Under Optimum Conditions

These data are not currently available.

4.3.4 PARTICULATE EMISSIONS FROM THE SMELT DISSOLVING TANK AND SLAKE TANK

The range of particulate emissions from the smelt dissolving tank and slake tank are of a low magnitude (1.0 - 4.0 lb/ADT and 4.0 - 6.0 lb/ADT respectively). The emissions are primarily caused by the entrainment of large particles in the vent gases. Because of the violent reactions taking place in each of these tanks, it is reasonable to expect that the turbulence of the dissolving water will splash water droplets containing both dissolved and undissolved inorganic salts above the surface. Here, because of the high temperature of the vent gases, the water may evaporate leaving the solid particles in suspension above the liquid. These particles may be carried out by the vent gases if they are not of sufficient weight to drop back into the liquid.

4.4 NSSC PROCESS - GENERAL

"As in the acid sulfite process, neutral sulfite pulping depends on sulphonation and hydrolysis of the lignin to soften the rigid matrix in which the wood fibers are bound together. Since sulphonation takes place slowly and to a limited extent, involving "A" groups of lignin only, high temperatures in the range of 160 to 190°C are required to complete the reaction within a reasonable time. Although the dissolution of the lignin is described as a two-stage process-sulphonation followed by hydrolysis-solubilization begins quite soon, since some of the lignin molecules are small enough to dissolve immediately on sulphonation, in contrast to the large molecules which must hydrolyze before they can pass into solution"(21).

"In semichemical pulping perhaps one-half or less of the total lignin is removed. In preparing bleachable pulps, it is not economical to cook much below a lignin content of 10 percent in the pulp because cooking time, carbohydrate loss, and chemical requirement increase disproportionately. Although the neutral character of the liquor is a handicap in sulphonation, it is beneficial insofar as it reduces loss of carbohydrate by hydrolysis. This contributes to the exceptionally high yield and strength of NSSC pulps"(21).

Because of the difference in the chemical attack on the lignin using sulfite liquors, such compounds as methyl mercaptan and dimethyl sulfide are not formed during digestion. The NSSC process should therefore be free from these odorous compounds. In addition, the absence of sulfide ions from the cooking liquor will virtually eliminate hydrogen sulfide as a possible emission.

NSSC pulp mills handle the spent cooking liquor in a variety of manners. Many NSSC mills are located on the same property as kraft mills. In these instances the spent cooking liquor can be mixed with the kraft black liquor and serve as make-up chemicals for the kraft process. A process similar to this is shown in the NSSC Flow Diagram No. 1 located in Chapter 5. It is possible to recover some of the chemicals from the smelt tank to provide fresh cooking liquor for the NSSC pulping although this is not shown.

Another method of handling the spent cooking liquor is to simply discharge the liquor to settling ponds, soil filtration or aeration units without an attempt to recover the chemicals. Such a system obviously will cause stream pollution. In the system shown in NSSC Flow Diagram No. 2, a percentage of the spent liquor is mixed with the fresh cooking liquor while the remaining spent liquor is discharged to the sewer.

For NSSC mills which do not have kraft recovery systems available, the chemicals of the spent liquor may be recovered in a fluidized bed reactor in the form of sodium sulfate and sodium carbonate. A fluidized bed recovery system is shown in NSSC Flow Diagram No. 3.

In each of the spent liquor handling systems mentioned above cooking liquor must be prepared from fresh chemicals. There are two methods available for the preparation of fresh liquor.

One is to mix soda ash and sodium sulfite together with the proper amount of water. Spent cooking liquor may be used to provide some of the chemicals and water.

Another method is to burn raw sulfur to sulfur dioxide in a sulfur burner and subsequently absorb the SO_2 in an absorption tower using a soda ash solution for an aqueous medium.

Atmospheric emission sources from an NSSC mill are limited to SO_2 absorption towers, blow pits, spent liquor evaporators, and fluidized bed reactors. In the case of spent liquor recovery in a kraft mill recovery system, the NSSC liquor will have an effect upon the emissions from the kraft recovery system. Each of these sources will be discussed in the following sections.

The quantities of atmospheric emissions from the NSSC emissions sources are unknown at the present time and subsequent discussions deal with qualitative aspects of emissions from the NSSC systems only.

4.4.1 GASEOUS EMISSIONS FROM THE NSSC PROCESSES

4.4.1.1 Sulfur Dioxide Absorption Tower

Sulfur dioxide absorption towers generally are countercurrent packed towers using sodium carbonate (soda ash) as the absorbing medium. The chemical absorption takes place according to the following reaction:

$$2\ Na^+ + CO_3 = +SO_2 \rightarrow 2\ Na^+ + SO_3^= + CO_2\uparrow \qquad \text{Eq. 4.23}$$

This reaction will lead to the emission of carbon dioxide. Sulfur dioxide may also be absorbed in water according to the following reaction:

$$HOH + SO_2 \rightarrow 2\ H^+ + SO_3^= \qquad \text{Eq. 4.24}$$

Nearly total absorption of sulfur dioxide in the tower is feasible in a properly designed and operated tower. The quantity of sulfur dioxide emitted from the tower will depend upon the efficiency of operation. Quantitative data of the emission of sulfur dioxide is therefore dependent upon the design and operating conditions of the individual towers.

4.4.1.2 Blow Pit

When the cooked pulp is blown into the blow pit, large amounts of steam and gases escape from the pulp and spent liquor. Sulfur dioxide seems to be the major gaseous emission from the blow pits. Recently, recovery systems have been installed to recover the sulfur dioxide from the blow pit gases. Installations in NSSC mills have been of individualistic design but all installations are based upon the absorption of sulfur dioxide by wet scrubbing either in the vent stack or by routing the gas through an SO_2 absorption tower. Again the efficiency of operation will dictate the quantity of sulfur dioxide emissions from the blow pit.

4.4.1.3 Fluidized Bed Reactor

Fluidized bed reactors have been developed for NSSC mills which do not have kraft mill recovery facilities for disposing of their spent liquors. Fluidized beds can accomplish the complete oxidation of organic wood constituents in the liquor yielding gaseous products of carbon dioxide, water, and minute traces of other organic compounds. Recovery of the cooking chemicals is in the form of sodium sulfate and sodium carbonate. These chemicals may be reused as kraft mill make-up chemicals. The ratio of sodium sulfate to sodium carbonate products may be controlled by the operating conditions in the fluidized bed.

In principle the spent cooking liquor is concentrated to approximately 30-40 percent solids and is sprayed into the top of the bed. As the liquor falls and comes into contact with the rising gases evaporation occurs. The temperature of the gases above the bed is approximately 800-1000°F. The partially dried liquor will either fall onto the bed or deposit on rising entrained dust particles and become thoroughly dried. As these particles grow they become too heavy to become entrained and will stay on the bed. Entrained particles which escape the reactor are collected in suitable mechanical dust collectors and are returned to the bed.

At the temperatures found above the bed, essentially none of the organic chemicals will undergo oxidation. The temperature of air supplied to the bed is approximately 1100-1200°F, while the bed temperature is in the range of 1200-1400°F, slightly higher than the air supplied because of the exothermic reactions taking place.

Under normal operating conditions atmospheric emissions from a fluidized bed reactor with dry and wet mechanical control devices are low. However, data on the actual emissions are not currently available.

4.4.1.4 Recovery in Kraft Systems

Precise data are not available on the effect which mixing spent NSSC liquor with kraft black liquor has upon the emissions from kraft recovery units from the point at which they are mixed up to and including the recovery furnace. However, the greatest effect that the mixing will have is to lower the pH of the kraft liquor. Section 4.2 details the importance of maintaining a high residual pH of the kraft black liquor during the recovery process. The effect of this mixing must be studied in greater detail.

4.5 EMISSIONS FROM THE SULFITE PROCESS

4.5.1 EMISSIONS FROM THE SULFITE PROCESS - GENERAL

There are four classes of sulfite pulping liquors: acid sulfite, bisulfite, neutral sulfite, and alkaline sulfite. Table 4-15 gives the predominant cooking chemical and initial pH of each of the classes. These two variables determine which class of sulfite pulping that a particular mill practices.

Calcium, sodium, magnesium, and ammonia are the four base chemicals around which the sulfite processes have been designed. At a pH below 6, it is proper to represent the sulfite in the cooking liquor as hydrosulfite ion (HSO_3^-) while above this pH it is represented as sulfite ion ($SO_3^=$). The calcium and sulfite combination is insoluble in aqueous solution of pH above 2. Hence, calcium sulfite cooking liquors are limited to the acid sulfite processes. Magnesium sulfite is soluble in solutions whose pH is below 7 (approximately), and it may be used in acid sulfite, bisulfite, and over the lower end of the neutral sulfite range of pH. Ammonium sulfite is soluble in solutions of a pH below 9 (approximately), while sodium sulfite is soluble over the entire range of pH. The desired range of pH for the cooking liquor will dictate the type of cooking chemicals which can be used.

As in the kraft pulping process, the objective of the sulfite cook is to dissolve the lignin in a wood chip and leave a fiber group which may be dispersed into an aqueous suspension suitable for paper making. The method of attack by sulfite liquors on lignin is different than the kraft liquor chemical attack. The sulfite and neutral sulfite processes involve lignin sulphonation, acid hydrolysis, and acid condensation (22). In sulfite cooking, the products of lignin-sulfite reactions do not produce volatile reduced sulfur compounds such as methyl mercaptan and dimethyl sulfide.

Sulfur dioxide is the principle atmospheric emission from the sulfite processes. The main causes of SO_2 release are stripping by gas streams and volatilization during periods of high liquor temperature. Hydrogen sulfide emissions are possible during recovery of the spent liquors if the recovery system is not maintained under proper oxidizing situations.

TABLE 4-15

PREDOMINATE CHEMICALS AND pH OF SULFITE COOKING LIQUORS (23)

Name	Predominate Chemical in Cooking Liquor	Approximate initial pH at 25°C
Acid sulfite	$H_2SO_3 + XHSO_3$	1 - 2
Bisulfite	$XHSO_3$	2 - 6
Neutral sulfite	$XSO_3 + XCO_3$	6 - 9+
Alkaline sulfite	$XSO_3 + XOH$	10+

4.5.2 GASEOUS EMISSIONS FROM SULFITE PROCESSES

4.5.2.1 Gaseous Emissions from Absorption Towers

Industrial absorption towers for sulfite processes are usually packed towers or Venturi absorbers. In the case of packed towers, sulfur dioxide gas is introduced in the bottom of the tower while a corbonate solution of the desired base (sodium, etc.) is introduced at the top of the tower. In the case of calcium, lime-rock ($CaCO_3$) is introduced as packing into the tower. Sulfur dioxide reacts with water to yield an acidic solution.

$$\underset{\xleftarrow{\text{basic}}}{\overset{\xrightarrow{\text{acidic}}}{2H^+ + SO_3^= \rightleftarrows H^+ + HSO_3^- \rightleftarrows H_2SO_3 \rightleftarrows H_2O + SO_2}} \qquad \text{Eq. 4.25}$$

Carbonate in solution will undergo an adjustment in equilibrium as the hydrogen ion concentration increases. The result is the production of carbon dioxide gas.

$$CO_3^= + H^+ \rightleftarrows CO_2 \uparrow + H_2O \qquad \text{Eq. 4.26}$$

This reaction depletes the concentration of hydrogen ion and tends to maintain a constant pH.

Acid fortification towers are absorption towers. Weak cooking liquor is passed through the tower for the purpose of absorbing additional sulfur dioxide. This replenishment of sulfite in the liquor offsets the sulfite lost through mill emissions as sulfur dioxide or combined in lignosulphonic acids in the pulping process.

Sufficient data are not available to adequately determine a range of sulfur dioxide emissions for gas absorption towers. The quantity of sulfur dioxide gas delivered to the absorption tower will depend upon the desired pH and strength of inorganic chemicals in the cooking liquor. Currently available data show that the sulfur dioxide emissions from the absorption tower are in the range of 15-20 lb/ADT before secondary scrubbing of the overhead gases.

4.5.2.2 Gaseous Emissions from Digester Relief and Blow Gases

Gaseous sulfur dioxide emissions from digester relief and blow gases stem from the temperature increase of the cooking liquor. Gases in general become less soluble in aqueous solutions during temperature rises. Sulfur dioxide is such a gas. Temperatures in the digester may range from 120°C to 180°C. When relief lines are opened and the pressure within the digester is relieved large quantities of sulfur dioxide will be emitted with the escaping steam.

There are three methods of discharging the digester; hot blowing, cold blowing, and flushing. In a hot blow, the pressure in the digester is relieved to a predetermined level and the contents are then blown into a blow pit.

In cold blowing, the pressure in the digester is relieved to a low level and the contents are then pumped into a dump tank below the digester. Spent liquor is introduced into the bottom cone of the digester to reduce pulp consistency and aid discharge.

In the flushing system, after the digester has been relieved, spent liquor or hot water is pumped into the digester for several minutes at a high rate. The blow valve is then opened and the pulp is discharged while the flushing liquid continues to enter the digester.

The three types of digester discharge affect the amount of sulfur dioxide which is emitted to the atmosphere. Gases which leave the digester during relief are sent to the accumulators where they fortify the cooking liquor. The gases which pass through the accumulators are sent to other areas in the process where they may be absorbed. However, blow tanks and dump tanks are usually vented to the atmosphere. Gases carried from the digester to these tanks may therefore be sources of emissions. Recent installation of gaseous control devices on blow pit gases will reduce these emissions.

A review of the digester discharge systems shows that for the hot blow style the pressure is only partially relieved before the blow is made. The digester gases which were not relieved will be sent to the blow pit during the blow. Significant quantities of sulfur dioxide are therefore emitted in this style of discharging if no recovery is practiced.

In the cold blow and flushing style of discharging the digester, the pressure is almost fully relieved, and the relief gases are routed to the accumulators. That fraction of the gases which remains in the digester may then escape from the system when the pulp is discharged to the dump tanks. Table 4-16 shows what ranges of sulfur dioxide emissions might be expected from the blow pit vent stack.

TABLE 4-16

APPROXIMATE EMISSIONS FROM BLOW PIT OR DUMP TANK VENTS
(WITHOUT SCRUBBING)

Blow Pit	100 - 150 lb/ADT
Dump Tank	10 - 25 lb/ADT

4.5.2.3 Gaseous Emissions During The Recovery of Spent Cooking Liquors

Practices in the recovery of the base used in pulping differs widely from mill to mill, as researchers find more effective methods of chemical reclamation. Because of the variety of chemical and physical properties exhibited by the base chemicals, calcium, sodium, magnesium, and ammonia, different processes have been developed to satisfy the handling and recovery problems peculiar to each base. In some instances no attempt is made to recover the chemical or sensible heat of the spent liquors, or in some cases only the heat is recovered.

The four sulfite processes presented in the flow diagrams located in Chapter 3 represent the four most widely used processes and the attendant recovery systems. They will be the recovery processes which will be discussed primarily in this section.

"In the case of calcium base liquors the scaling of evaporator surfaces, the need for all stainless-steel equipment, the relatively low value and the chemical complexity of the resultant smelt, and the difficulty with fly ash" (24) have led to the common practice of simply disposing of the spent liquors by a convenient means and making no attempt to recover the heat or chemicals. Sulfite mills which incorporate these processes usually utilize sulfur dioxide absorption towers in conjunction with a sulfur burner as shown in Sulfite Flow Diagram No. 2. The digester relief gases also may be routed through the accumulator tanks for recovery of sulfur dioxide. Gases vented to the accumulators are further routed through the acid-making system to recover as much SO_2 as possible. Spent liquors which are disposed of by means of a sewer may present severe water pollution hazards.

Spent "liquor from several magnesium sulfite processes can be burned in a simple heat- and chemical-recovery system in which the inorganic salts break down into magnesium oxide and sulfur dioxide. These chemicals can then be recombined directly to produced magnesium bisulfite acid for cooking" (25). Sulfite Flow Diagrams No. 1 and 3 show variations of such a process. A thorough discussion of the recovery process principles may be found in "The Pulping of Wood" (25).

The sulfur dioxide produced in the recovery furnace is carried out and through a dust collector before entering the direct contact evaporator. These evaporators are designed to prevent absorption of the SO_2 in the evaporating liquor, so that all SO_2 may be sent to the absorption tower.

The venturi absorption towers absorb the sulfur dioxide with a solution of magnesium hydroxide. Sulfur dioxide enters the towers from the recovery furnace as well as from the fortification towers, and/or the digester. Magnesium hydroxide readily absorbs the sulfur dioxide. Absorption efficiencies of the venturi absorption systems range from 95 - 98+ percent. Sulfur dioxide emissions from the absorption system range from 10 - 25 lb/ADT of pulp.

"At present there is no process in commercial operation for the recovery of pulping chemicals from ammonium-base spent cooking liquors" (26). Many ammonium acid sulfite mills simply incinerate their waste liquors in combination boilers to recover the heating value of the liquor. Such a system is demonstrated in Sulfite Flow Diagram No. 4. Sulfur dioxide emissions from such incinerators are in the range of 250 - 500 lb/ADT.

Sulfur trioxide emissions from recovery furnaces and incinerators are possible if close control is not taken over the excess oxygen in the flue gases as well as its temperature. A common means for controlling SO_3 formation, other than oxygen control, is the use of cooling towers which cool the sulfur dioxide below the temperature required for conversion to sulfur trioxide. Quantitative data of the emission of sulfur trioxide are not available.

4.5.2.4 Emissions from Multiple-Effect Evaporators

Multiple-effect evaporators which concentrate spent liquors are a source of sulfur dioxide emissions. Such emissions are evolved because of the high temperature and low pressure conditions in the effects. The type of condenser has an effect on the sulfur dioxide emissions. Adequate contact between the sulfur dioxide and the cooling water will remove a large portion of the gas.

Sulfur dioxide emissions from the multiple effect evaporators are in the range of 5 - 10 lb/ADT of pulp.

4.6 AUXILIARY FURNACES

The atmospheric emissions of sulfur dioxide and fly ash from auxiliary furnaces are directly dependent upon the amount of sulfur bearing fuels which are burned in the furnace. The amount of energy which must be provided by the auxiliary furnaces is directly dependent upon the total energy requirements of a mill and the amount of this energy which can be furnished by the recovery system. This energy demand on the auxiliary furnace is highly individualistic for each pulp and paper mill.

A common source of fuel for auxiliary furnaces is wood bark. Bark has a heating value of approximately 4,500 BTU/lb., and is readily available in mills which use roundwood. Bark burning produces very little sulfur dioxide; approximately 1 lb.SO_2/1000 lbs. bark. Fly ash from bark burning is in the range of approximately 10 - 20 lb. ash/1000 lbs. bark.

Fuel oil is another common fuel used in auxiliary furnaces. Fuel oil contains varying percentages of sulfur which forms sulfur dioxide in the oxidizing atmosphere of the furnace. Fuel oil has a heating value of approximately 19,000 BTU/lb. The quantity of sulfur dioxide formed in the combustion of fuel oil, assuming efficient combustion, depends mainly upon the sulfur content of the oil, which is usually in the range of 1 - 6 percent. A convenient formula for determining the amount of sulfur dioxide produced in the combustion of fuel oil is:

$$1000 \; \frac{\text{pound } SO_2 \text{ formed}}{\text{pounds oil burned}} = \frac{\text{pounds sulfur}}{\text{pound fuel oil}} \times \frac{\text{2 pounds } SO_2}{\text{pound sulfur}} \times 1000$$

The fly ash formation of fuel oil ranges from 0.9 - 1.0 lbs. ash/1000 lbs oil.

Coal is another fuel used in auxiliary furnaces. Coal has a heating value of 14,000 BTU/lb. Like fuel oil, the sulfur content of coal is usually in the range of 1 - 6 percent. The quantity of sulfur dioxide formed in the combustion of coal may be found by using the equation shown above. The fly ash produced by burning coal is higher than that of fuel oil, and is in the range of 60 - 80 lbs/1000 lb coal.

Natural gas is another fuel which is now being used to a limited extent for auxiliary furnaces. Its heating value is 1,000 BTU/ft^3. The sulfur and ash content of natural gas may be considered negligible.

Table 4-17 summarizes the data.

TABLE 4-17

SULFUR DIOXIDE AND FLY ASH PRODUCTION FROM THE COMBUSTION OF ASSORTED FUELS IN AN AUXILIARY FURNACE

	SO_2 (lb/1000 lbs fuel)	Fly Ash (lb/1000 lbs fuel)
Bark	1.0 - 1.5	10 - 20
Oil (2% S)	40	0.9 - 1.0
Coal (2% S)	40	60 - 80
Natural Gas	Trace	Trace

Because of the varied power requirements placed upon the auxiliary furnaces by the assorted pulping and paper making systems, it is beyond the scope of this chapter to attempt to give emissions of sulfur dioxide and fly ash from the auxiliary furnaces in terms of pounds per air dry ton of pulp. However the flow diagrams and the associated heat and energy balances presented in Chapter 3, may give the reader an idea of the energy requirements placed on the auxiliary furnace. The flow diagrams show how these requirements can be satisfied using the fuels discussed here, and the emissions of sulfur dioxide and fly ash which might result.

4.7 SUMMARY OF EMISSION DATA

Table 4-18 summarizes the approximate ranges of emissions for the principle gaseous sulfur compound and particulate emissions from the kraft pulping and recovery system. Table 4-19 summarizes the approximate ranges of sulfur dioxide and fly ash formation in the combustion of fuels in the auxiliary furnaces.

Because of the variety and complexity of the pulping and recovery systems used in the sulfite and NSSC processes, a meaningful table cannot be presented here. The reader is referred to the flow diagrams in Chapter Three to obtain the emissions of a particular system.

TABLE 4-18

SUMMARY OF EMISSION DATA - KRAFT PROCESS

(Emissions in lb/ADT)

	SO_2	H_2S	MeSH	MeSMe	MeSSMe	Particulate
Digester Relief & Blow	T - 0.01	0.01 - 0.12	0.02 - 0.40	0.40 - 2.5	0.20 - 1.50	-
Brown Stock Washers	0.01 - 0.02	0.01 - 0.12	0.10 - 0.25	0.01 - 0.02	0.01 - 0.02	-
Oxidation Towers	0 - 0.01	0.01 - 0.02	0.05 - 0.10	0.02 - 0.08	0.05 - 0.15	-
Multiple Effect Evaporators OX.	0.01	0.01 - 0.02	0.10 - 0.30	0.05 - 0.15	0.05 - 0.15	-
UNOX.	0 - 0.01	0.10 - 3.0	0.10 - 1.50	0.05 - 0.08	0.01 - 0.02	-
Direct Contact Evaporators OX.	2.0 - 8.0	0.10 - 2.0	0.05 - 0.25	0.01 - 0.10	0.01 - 0.20	75-125(cascade) 20-40 (venturi)
UNOX.	2.0 - 8.0	5.0 - 30.0	0.50 - 2.50	0.10 - 0.30	0.10 - 0.40	75-125(cascade) 20-40 (venturi)
Recovery Furnace	10 - 15	1.0 - 5.0	0.01 - 0.10	0.01 - 0.02	0.01 - 0.02	
Smelt Tank	0.0 - 0.01	0.02 - 0.05	0.02 - 0.05	0.01 - 0.02	0 - 0.01	1.0 - 4.0
Lime Kiln	DATA NOT AVAILABLE - TRS - 0.01 - 0.83					20 - 65

TABLE 4-19

SUMMARY OF SULFUR DIOXIDE AND FLY ASH FORMATION IN AUXILIARY FURNACE COMBUSTION

	SO_2 (lb/1000 lb fuel)	Fly Ash (lb/1000 lb fuel)
Bark	1.0 - 1.5	10 - 20
Fuel Oil (2% sulfur)	40	0.9 - 1.0
Coal (2% sulfur)	40	60 - 80
Natural Gas	Trace	Trace

4.8 REVIEW OF EMISSION STANDARDS APPLICABLE TO PULP MILLS

A variety of emission regulations is in effect in many city, county, regional, and state abatement areas in the U.S. In addition to general nuisance provisions, specific emission limitations are in effect on particulates, individual gaseous compounds, and classes of compounds. Some of these limitations, such as emissions of particulates, sulfur dioxide, hydrogen sulfide, and total reduced sulfur compounds (TRS) may be applied to chemical pulp mills.

In some localities an attempt is made to apply to pulp mills the process weight table originally developed in Los Angeles County. This is an inappropriate application of the process weight table which was developed for and applied to the metallurgical industry. The process weight concept represents the state of the art on emissions control from metallurgical processes. The nature and size range of particulates, as well as the characteristics of the processes themselves, are vastly different from the recovery and calcining operations involved in chemical pulping.

As of late 1969, in states where pulp mills are located, regulations specific to pulp mills were to be found in Washington, Oregon, and Humboldt County, California. In all cases the rules are applicable to kraft mills only.

Provisions of the Washington and Oregon regulations are identical with respect to emission limitations. All limitations are stated per ton of air dry pulp produced. Principal provisions of the regulations include:

(a) TRS Compounds from the recovery furnace: No more than 2 pounds per ton (1972) reduced to no more than 1/2 pound per ton by 1975.

(b) Noncondensible gases from the digesters and multiple effect evaporators: Collected and burned in the lime kiln or proven equivalent.

(c) Particulates from the recovery furnace: No more than 4 pounds per ton.

(d) Particulates from lime kiln: No more than 1 pound per ton.

(e) Particulates from smelt tank: No more than 1/2 pound per ton.

The principal provisions of the Humboldt County regulations specific to pulp mills include:

(a) TRS compounds from any single emission point: No more than the total daily weight calculated by the formula:

$$\text{TRS (pounds per day)} = 0.012\ (H_s)^2$$

where H_s is the height in feet of the emission point above mean ground elevation.

(b) Total maximum allowable monthly TRS emissions to the atmosphere: No more than one pound per ton of dry wood charged into the [pulping] process.

(c) Maximum allowable ground level concentration of TRS, expressed as H_2S: No more than 0.03 ppm for no longer than 60 minutes.

In Washington and Oregon (29) the regulations were developed in cooperation with representatives of the kraft pulping industry. The most important sources of odors and particulates were identified, the state of the art in emissions control was determined, and developments in new technology were assessed. The regulations consider best available technology and uniqueness of process.

The states of Washington and Oregon presently are developing regulations for atmospheric emissions from sulfite mills.

By mid-1970, Air Quality Regions called for by the Air Quality Act of 1967 will be designated in all of the states. The criteria and control documents for SO_2 and particulates have been published by NAPCA. Thus more stringent SO_2 and particulate emission standards would appear inevitable under this program. Present plans by NAPCA call for publication of some 35 criteria documents (including one for odors) by 1975. Application of emission standards in the states should follow within about 15 months of publication.

4.9 REFERENCES

1. Cederlöf, R., Edfor, M. L., Friberg, L., Lindvall, T., Nordisk Hygenish Tidskrfit 46, 51, (1965).

2. Leonardis, G., Kendall, D., Barnard, N., "Odor Threshold Determinators of 53 Odorant Chemicals" J. APCA 19 (2), 91-5, (1969).

3. Young, F. A., Adams, D. R., Sullivan, Dobbs, "The Relationship between Environmental-Demographic Variables and Olfactory Detection and Objectionability Thresholds" to be published in Perception and Pschophysics.

4. "Handbook of Air Pollution" U. S. Department of H.E.W., P.H.S., Bureau of State Services, Division of Air Pollution, Cincinnati, Ohio.

5. "The Merck Index" 8th Edition, Merck and Co., Inc., Rahway, N. J., 1968.

6. Douglass, I. B., "Some Chemical Aspects of Kraft Odor Control", J. APCA 18 (8), 543, (1968)

7. McKean, W. R., Hrutfiord, B. F., Sarkanen, K. V., "Effect of Kraft Pulping Conditions on the Formation of Methyl Mercaptan and Dimethyl Sulfide," TAPPI, 50 (8), 400-05, (1968)

8. McKean, W. R., Hrutfiord, B. F., Sarkanen, K. V., "Kinetic Analysis of Odor Formation in the Kraft Pulping Process," TAPPI, 48 (12), 699-703, (1965).

9. Shih, T. T., Hrutfiord, B. F., Sarkanen, K. V., Johanson, L. N., "Methyl Mercaptan Vapor-Liquid Equilibrium in Aqueous Systems as a Function of Temperature and pH," TAPPI, 50 (12), 634-8, (1967).

10. Murray, F. E., Rayner, H. B., "The Emission of Hydrogen Sulfide from Kraft Recovery Furnaces," Pulp and Paper Magazine of Canada, 69 (5), 71-4, (1968).

11. Blosser, R. O., Cooper, H. B., Megy, J. A., Duncan, L., Tucker, T. W., "Factors Affecting Gaseous Sulfur Emissions in the Kraft Recovery Furnace Complex," Paper Trade Journal 153 (21), 58-59, (1969).

12. Harding, C. I., Landry, J. O., "Future Trends in Air Pollution Control in the Kraft Pulping Industry," *TAPPI*, 49 (8), 61-7a, (1966).

13. Thoen, G. N., DeHaas, G. G., Tallent, R. G., Davis, A. S., "The Effect of Combustion Variables on the Release of Odorous Sulfur Compounds from a Kraft Recovery Unit," *TAPPI*, 51 (8), 329-33, (1968).

14. Murray, F. E., Rayner, H. B., "Emission of Hydrogen Sulfide from Kraft Black Liquor during Direct-Contact Evaporation," *TAPPI*, 48 (10), 588-92, (1965).

15. *"The Pulping of Wood,"* MacDonald, R. G., Volume I, Second Edition, McGraw Hill Book Company, New York, (1969).

16. Ibid, page 455.

17. Ibid, page 478.

18. McKean, W. T., Hrutfiord, B. F., Sarkanen, K. V., "Kinetics of Methyl Mercaptan and Dimethyl Sulfide Formation in Kraft Pulping," *TAPPI*, 51 (12), 564-7, (1968).

19. Taylor, C. E., "Lime Kilns and Their Operation," in *Atmospheric Emissions from Sulfate Pulping*, (E. R. Hendrickson, Editor), April 1966.

20. Harding, C. I., "Source Reduction in the Pulping Industry," presented at the Fifth Annual Sanitary and Water Resources Engineering Conference, Vanderbilt University, June 2-3, 1966.

21. *"The Pulping of Wood,"* MacDonald, R. G., Volume I, Second Edition, McGraw Hill Book Company, New York, 1969, page 227-29.

22. Ibid, page 61.

23. Ibid, page 278.

24. Ibid, page 324-5

25. Ibid, page 341-3.

26. Ibid, page 341.

27. Walther, J. F., Amberg, H. R., "The Status of Odor Control in the Kraft Pulp Industry," presented at National AIChE Meeting, Portland, August 1969.

28. Shih, T. T., Hrutfiord, B. F., Sarkanen, K. V., Johanson, L. N., "Hydrogen Sulfide Vapor-Liquid Equilibrium in Aqueous Systems As A Function of Temperature and pH," TAPPI, 50 (12), 630-4, (1967).

29. Hildebrandt, P. W., Droege, H., and Stockman, R. L., "Development of Regulations for Atmospheric Emissions from Kraft Pulp Mills," presented at National AIChE Meeting, Atlanta, February 1970.

APPENDIX A

Appendix A consists of seven tables. Table A-1 summarizes the number of pulp mills and pulp production by process for each state.

Table A-2 presents specific data for each mill within a given state. The pulp tonnage listed in the columns headed Unbleached (Unbl.) is the total tonnage of all air dry pulp produced at a mill by a particular pulping process. The tonnage listed in the columns headed Bleached (Bl.) is that portion of the total tonnage which is bleached and is expressed in finished bleached tons. For example:

			Kraft	
City	County	Owner	Unbl.	Bl.
Ashdown	Little River	Nekoosa-Edwards Paper Company	430	400

means that the entire output of 430 unbleached tons was probably used to produce 400 finished bleached tons. Since a shrinkage (pulp loss) of approximately 5 - 10 percent is experienced from unbleached to finished bleached pulp, this should be a reasonable assumption

Jesup	Wayne	ITT, Rayonier	675	675

means that the mill reported the same tonnage for unbleached and bleached. These tonnages have been tabulated as received from the mill; whereas, in actual practice there should be a difference between the two numbers.

Mill age has not been included in these tabulations since many of the older mills have been modified to such an extent that the original mill age has little relation to the technological age of the mill.

Table A-3 contains detailed historical data presented in support of Figure 2-3.

Tables A-4 through A-7 present detailed data in support of Table 2-3.

T A B L E A-1

SUMMARY OF CHEMICAL WOOD PULP MILLS IN U. S.
AS OF DECEMBER 1968

LOCATION	NUMBER OF PULP MILLS AND PULP CAPACITY IN AIR DRY TONS PER DAY							
	KRAFT		SULFITE		SODA		NSSC	
STATE	NO.	AD TPD	NO.	AD TPD	NO.	AD TPD	NO.	AD TPD
Alabama	12	9,470	-	-	-	-	-	-
Alaska	-	-	2	1,340	-	-	-	-
Arizona	1	320	-	-	-	-	-	-
Arkansas	7	3,820	-	-	-	-	-	-
California	5	2,015	-	-	-	-	-	-
Colorado	-	-	-	-	-	-	-	-
Connecticut	-	-	-	-	-	-	-	-
Delaware	-	-	-	-	-	-	-	-
District of Columbia	-	-	-	-	-	-	-	-
Florida	9	8,710	1	375	-	-	-	-
Georgia	11	12,235	-	-	-	-	2	600
Hawaii	-	-	-	-	-	-	-	-
Idaho	1	840	-	-	-	-	-	-
Illinois	-	-	-	-	-	-	-	-
Indiana	-	-	-	-	-	-	2	370
Iowa	-	-	-	-	-	-	1	110
Kansas	-	-	-	-	-	-	-	-
Kentucky	-	-	-	-	-	-	1	200
Louisiana	9	7,966	-	-	-	-	4	915
Maine	6	2,505	4	1,290	-	-	2	360
Maryland	1	740	-	-	-	-	-	-
Massachusetts	-	-	-	-	1	54	-	-
Michigan	2	440	1	168	-	-	3	770
Minnesota	2	575	1	135	-	-	2	360
Mississippi	4	4,405	-	-	-	-	-	-
Missouri	-	-	-	-	-	-	-	-
Montana	1	1,200	-	-	-	-	-	-

TABLE A-1 (Continued)

STATE	KRAFT		SULFITE		SODA		NSSC	
	NO.	AD TPD	NO.	AD TPD	NO.	AD TPD	NO.	AD TPD
Nebraska	-	-	-	-	-	-	-	-
Nevada	-	-	-	-	-	-	-	-
New Hampshire	1	615	2	205	-	-	2	445
New Jersey	-	-	-	-	-	-	-	-
New Mexico	-	-	-	-	-	-	-	-
New York	1	205	2	160	1	155	4	611
North Carolina	4	4,440	-	-	-	-	2	570
North Dakota	-	-	-	-	-	-	-	-
Ohio	1	652	-	-	-	-	2	790
Oklahoma	-	-	-	-	-	-	1	420
Oregon	6	4,745	5	765	-	-	1	250
Pennsylvania	5	1,235	2	350	1	86	1	385
Rhode Island	-	-	-	-	-	-	-	-
South Carolina	4	4,860	-	-	-	-	2	725
South Dakota	-	-	-	-	-	-	-	-
Tennessee	2	1,220	-	-	1	275	2	309
Texas	5	3,780	-	-	-	-	-	-
Utah	-	-	-	-	-	-	-	-
Vermont	-	-	-	-	-	-	-	-
Virginia	4	4,000	-	-	-	-	4	1,110
Washington	8	5,560	12	4,515	-	-	3	510
West Virginia	-	-	-	-	-	-	-	-
Wisconsin	4	1,255	11	1,552	-	-	2	865
Wyoming	-	-	-	-	-	-	-	-
TOTAL	116	87,808	43	10,875	4	570	43	10,675

TABLE A-2

SUMMARY OF CHEMICAL WOOD PULP MILLS IN EACH STATE AS OF DECEMBER 1968

LOCATION			KIND OF PULP MILLS AND AD TPD CAPACITY								TYPE WOOD	
			KRAFT		SULFITE		SODA		NSSC		SOFT-WOOD	HARD-WOOD
CITY	COUNTY	OWNER	Unbl.	Bl.	Unbl.	Bl.	Unbl.	Bl.	Unbl.	Bl.	%	%
			ALABAMA									
Brewton	Escambia	Container Corp. of America	900	400							75	25
Coosa Pines	Talladega	Kimberly-Clark Corp.	630	630							64	36
Demopolis	Marengo	Gulf States Paper Corp.	400	375							60	40
Jackson	Clarke	Allied Paper, Inc.	520	475							40	60
Mahrt	Russell	Georgia Kraft Co.	900	0							90	10
Mobile	Mobile	International Paper Co., Sou. Kraft Div.	1050	450							77	23
Mobile	Mobile	Scott Paper Co.	1400	1300							73	27
Montgomery	Autauga	Union Camp Corp.	870	0							93	7
Naheola	Choctaw	American Can Co.	970	900							60	40
Selma	Dallas	Hammermill Paper Co.	430	400							75	25
Pine Hill	Wilcox	MacMillan Bloedel United, Inc.	900	0							95	5

T A B L E A-2 (Continued)

CITY	COUNTY	OWNER	KRAFT Unbl.	KRAFT Bl.	SULFITE Unbl.	SULFITE Bl.	SODA Unbl.	SODA Bl.	NSSC Unbl.	NSSC Bl.	SOFT-WOOD %	HARD-WOOD %
Tuscaloosa	Tuscaloosa	Gulf States Paper Corp.	500	0							90	10
TOTAL ALABAMA		Unbleached =	9470		0		0		0			
		Of Which Bleached =		4930								
		NUMBER OF MILLS		12		0		0		0		
ALASKA												
Ketchikan		Ketchikan Pulp Co.			740	620					100	0
Sitka		Alaska Lumber & Pulp Co., Inc.			600	500					100	0
TOTAL ALASKA		Unbleached =	0		1340		0		0			
		Of Which Bleached =				1120						
		NUMBER OF MILLS		0		2		0		0		

T A B L E A-2 (Continued)

CITY	COUNTY	OWNER		KRAFT		SULFITE		SODA		NSSC		SOFT-WOOD	HARD-WOOD
				Unbl.	Bl.	Unbl.	Bl.	Unbl.	Bl.	Unbl.	Bl.	%	%
					ARIZONA								
Snowflake	Navajo	Southwest Forest Industries, Inc.		320	65							100	0
TOTAL ARIZONA		Unbleached	=	320		0		0		0			
		Of Which Bleached	=		65								
		NUMBER OF MILLS		1		0		0		0			
					ARKANSAS								
Ashdown	Little River	Nekoosa-Edwards Paper Co.		430	400							65	35
Camden	Ouachita	International Paper Co., Sou. Kraft Div.		700	0							93	7
Crossett	Ashley	Georgia-Pacific Corp., #1 Mill		640	120							100	0
		#2 Mill		200	200							0	100
Morrilton	Conway	Arkansas Kraft Corp.		350	0							88	12
Pine Bluff	Jefferson	Dierks Paper Co.		200	0							100	0
Pine Bluff	Jefferson	International Paper Co., Sou. Kraft Div.		1300	1170							70	30
TOTAL ARKANSAS		Unbleached	=	3820		0		0		0			
		Of Which Bleached	=		1890								
		NUMBER OF MILLS		7			0		0		0		

T A B L E A-2 (Continued)

CITY	COUNTY	OWNER	KRAFT		SULFITE		SODA		NSSC		SOFT-WOOD	HARD-WOOD
			Unbl.	Bl.	Unbl.	Bl.	Unbl.	Bl.	Unbl.	Bl.	%	%
			CALIFORNIA									
Anderson	Shasta	Kimberly-Clark Corp.	165	150							100	0
Antioch	Contra Costa	Fibreboard Corp.	500	160							100	0
Antioch	Contra Costa	Fibreboard Corp.	250	0							85	15
Fairhaven	Humboldt	Crown Simpson Pulp Co.	550	500							100	0
Samoa	Humboldt	Georgia-Pacific Corp.	550	500							100	0
TOTAL CALIFORNIA		Unbleached =	2015		0		0		0			
		Of Which Bleached =		1310								
		NUMBER OF MILLS		5		0		0		0		
			FLORIDA									
Fernandina Beach	Nassau	Container Corp. of America	825	0							95	5
Fernandina	Nassau	ITT Rayonier			375	375					100	0
Foley	Taylor	The Buckeye Cellulose Corp.	950	860							85	15

T A B L E A-2 (Continued)

CITY	COUNTY	OWNER		KRAFT Unbl.	KRAFT Bl.	SULFITE Unbl.	SULFITE Bl.	SODA Unbl.	SODA Bl.	NSSC Unbl.	NSSC Bl.	SOFT-WOOD %	HARD-WOOD %
		FLORIDA (Continued)											
Jacksonville	Duval	Alton Box Board Co.		675	0							90	10
Jacksonville	Duval	St. Regis Paper Co.		1300	0							60	40
Palatka	Putnam	Hudson Pulp & Paper Corp.		950	400							60	40
Panama City	Bay	International Paper Co., Sou. Kraft Div.		1410	735							50	50
Pensacola	Escambia	St. Regis Paper Co. - West Mill		300	260							88	12
Port St. Joe	Gulf	St. Joe Paper Co.		1700	500							90	10
TOTAL FLORIDA		Unbleached	=	8710		375		0		0			
		Of Which Bleached	=		2755		375						
		NUMBER OF MILLS		9		1			0		0		
		GEORGIA											
Augusta	Richmond	Continental Can Co., Inc.		750	700							60	40
Brunswick	Glynn	Brunswick Pulp & Paper Co.		1290	1195							61	39

T A B L E A-2 (Continued)

CITY	COUNTY	OWNER	KRAFT Unbl.	KRAFT Bl.	SULFITE Unbl.	SULFITE Bl.	SODA Unbl.	SODA Bl.	NSSC Unbl.	NSSC Bl.	SOFT-WOOD %	HARD-WOOD %
			GEORGIA (Continued)									
Cedar Springs	Early	Great Northern Paper Co.	1700	0							99	1
Cedar Springs	Early	Great Northern Paper Co.							300	0	0	100
Jesup	Wayne	ITT Rayonier, Inc.	675	675							90	10
Macon	Bibb	Georgia Kraft Co.	875	0							85	15
Port Wentworth	Chatham	Continental Can Co., Inc.	625	0							95	5
Riceboro	Liberty	Interstate Paper Corp.	450	0							95	5
Rome	Floyd	Georgia Kraft Co.	1500	0							90	10
St. Marys	Camden	Gilman Paper Co.	900	350							80	20
Savannah	Chatham	Union Camp Corp.	2700	0							95	5
Savannah	Chatham	Union Camp Corp.							300	0	0	100
Valdosta	Lowndes	Owens-Illinois Inc.	770	0							95	5
TOTAL GEORGIA		Unbleached =	12,235		0		0		600			
		Of Which Bleached =		2,920						0		
		NUMBER OF MILLS	11			0		0	2			

T A B L E A-2 (Continued)

CITY	COUNTY	OWNER		KRAFT Unbl.	KRAFT Bl.	SULFITE Unbl.	SULFITE Bl.	SODA Unbl.	SODA Bl.	NSSC Unbl.	NSSC Bl.	SOFT-WOOD %	HARD-WOOD %
					IDAHO								
Lewiston	Nez Perce	Potlatch Forests, Inc.		840	800							100	0
TOTAL IDAHO		Unbleached	=	840		0		0		0			
		Of Which Bleached	=		800								
		NUMBER OF MILLS		1		0		0		0			
					INDIANA								
Carthage	Rush	Container Corp. of America								120	0	0	100
Terre Haute	Vigo	Weston Paper & Mfg. Co.								250	0	0	100
TOTAL INDIANA		Unbleached	=	0		0		0		370			
		Of Which Bleached	=								0		
		NUMBER OF MILLS		0		0		0		2			

T A B L E A-2 (Continued)

CITY	COUNTY	OWNER		KRAFT Unbl.	KRAFT Bl.	SULFITE Unbl.	SULFITE Bl.	SODA Unbl.	SODA Bl.	NSSC Unbl.	NSSC Bl.	SOFT-WOOD %	HARD-WOOD %
						IOWA							
Fort Madison	Lee	Consolidated Packaging Corp.								110	0	0	100
TOTAL IOWA		Unbleached	=	0		0		0		110			
		Of Which Bleached	=								0		
		NUMBER OF MILLS		0		0		0		1			
						KENTUCKY							
Hawesville	Hancock	WesCor Corp.								200	0	0	100
TOTAL KENTUCKY		Unbleached	=	0		0		0		200			
		Of Which Bleached	=								0		
		NUMBER OF MILLS		0		0		0		1			

TABLE A-2 (Continued)

CITY	COUNTY (Parish)	OWNER	KRAFT		SULFITE		SODA		NSSC		SOFT-WOOD	HARD-WOOD
			Unbl.	Bl.	Unbl.	Bl.	Unbl.	Bl.	Unbl.	Bl.	%	%
		LOUISIANA										
Bastrop	Morehouse	International Paper Co., Sou. Kraft Div. Bastrop Mill							485	0	0	100
Bastrop	Morehouse	International Paper Co., Sou. Kraft Div. Louisiana Mill	1200	1100							45	55
Bogalusa	Washington	Crown Zellerbach Corp.	1350	140							99	1
Bogalusa	Washington	Crown Zellerbach Corp.							150	0	0	100
Elizabeth	Allen	Calcasieu Paper Co., Inc.	250	0							90	10
Hodge	Jackson	Continental Can Co., Inc.	500	0							100	0
Hodge	Jackson	Continental Can							200	0	0	100
Pineville	Rapides	Pineville Kraft Corp.	850	0							100	0
Port Hudson	E. Baton Rouge	Louisiana Forest Products Corp.	550	510							15	85
St. Francisville	West Feliciana	Crown Zellerbach Corp.	550	500							60	40
Springhill	Webster	International Paper Co., Sou. Kraft Div.	1650	1000							70	30
W. Monroe	Ouachita	Olinkraft, Inc.	1066	0							77	23

TABLE A-2 (Continued)

CITY	COUNTY	OWNER	KRAFT		SULFITE		SODA		NSSC		SOFT-WOOD	HARD-WOOD
			Unbl.	Bl.	Unbl.	Bl.	Unbl.	Bl.	Unbl.	Bl.	%	%
			LOUISIANA (Continued)									
W. Monroe	Ouachita	Olinkraft, Inc.	1066	0							77	23
W. Monroe	Ouachita	Olinkraft, Inc.							80	0	0	100
TOTAL LOUISIANA		Unbleached =	7966						915			
		Of Which Bleached =		3250						0		
		NUMBER OF MILLS	9		0		0		4			
			MAINE									
Augusta	Kennebec	Statler Tissue Corp.			125	125					85	15
Cumberland Mills	Cumberland	S. D. Warren Co.	270	250							30	70
East Millinocket	Penobscot	Great Northern Paper Co.							160	0	0	100
Jay	Franklin	International Paper Co.	525	470							65	35
Lincoln	Penobscot	Lincoln Pulp	225	210							0	100

T A B L E A-2 (Continued)

CITY	COUNTY	OWNER		KRAFT Unbl.	KRAFT Bl.	SULFITE Unbl.	SULFITE Bl.	SODA Unbl.	SODA Bl.	NSSC Unbl.	NSSC Bl.	SOFT-WOOD %	HARD-WOOD %
				MAINE (Continued)									
Millinocket	Penobscot	Great Northern				500	0					100	0
Old Town	Penobscot	Penobscot Co.		375	350							10	90
Old Town	Penobscot	Penobscot Co.				215	200					100	0
Rumford	Oxford	Oxford Paper Co.		560	525							35	65
Winslow	Kennebec	Scott Paper Co.				450	415					75	25
Woodland	Washington	Georgia-Pacific Corp.		550	500							70	30
Woodland	Washington	Georgia-Pacific Corp.								200	0	0	100
TOTAL MAINE		Unbleached	=	2505		1290		0		360			
		Of Which Bleached	=		2305		740				0		
		NUMBER OF MILLS		6		4			0	2			
				MARYLAND									
Luke	Allegany	West Virginia Pulp & Paper Co.		740	740							30	70
TOTAL MARYLAND		Unbleached	=	740		0		0		0			
		Of Which Bleached	=		740								
		NUMBER OF MILLS		1			0		0		0		

T A B L E A-2 (Continued)

CITY	COUNTY	OWNER		KRAFT Unbl.	KRAFT Bl.	SULFITE Unbl.	SULFITE Bl.	SODA Unbl.	SODA Bl.	NSSC Unbl.	NSSC Bl.	SOFT WOOD %	HARD-WOOD %
						MASSACHUSETTS							
Lawrence	Essex	Oxford Paper Co.						54	50			0	100
TOTAL MASSACHUSETTS		Unbleached	=	0		0		54		0			
		Of Which Bleached	=						50				
		NUMBER OF MILLS			0		0	1			0		
						MICHIGAN							
Detroit	Wayne	Scott Paper Co.*				168	168					90	10
Filer City	Manistee	Packaging Corp. of America		200	184							50	50
Filer City	Manistee	Packaging Corp. of America								375	0	0	100
Muskegon	Muskegon	S. D. Warren Co.		240	225							36	64
Ontonagon	Ontonagon	Hoerner Waldorf Corp.								225	0	0	100
Ostego	Allegan	Menasha								170	0	0	100
TOTAL MICHIGAN		Unbleached	=	440		168		0		770			
		Of Which Bleached	=		409		168				0		
		NUMBER OF MILLS		2		1		0		3			

*Shut down in 1969

T A B L E A-2 (Continued)

CITY	COUNTY	OWNER	KRAFT		SULFITE		SODA		NSSC		SOFT-WOOD	HARD-WOOD
			Unbl.	Bl.	Unbl.	Bl.	Unbl.	Bl.	Unbl.	Bl.	%	%
			MINNESOTA									
Cloquet	Carlton	The Northwest Paper Company Sub. of Potlatch Forests, Inc.	325	305							60	40
Cloquet	Carlton	The Northwest Paper Company Sub. of Potlatch Forests, Inc.			135	125					10	90
Grand Rapids	Itasca	Blandin Paper Co.							50	50	0	100
International Falls	Koochiching	Boise Cascade Corp.	250	250							50	50
St. Paul	Ramsey	Hoerner Waldorf Corp.							310	0	0	100
TOTAL MINNESOTA		Unbleached =	575		135		0		360			
		Of Which Bleached =		555		125		0		50		
		NUMBER OF MILLS	2		1		0		2			
			MISSISSIPPI									
Monticello	Lawrence	St. Regis Paper	1690	0							100	0

T A B L E A-2 (Continued)

CITY	COUNTY	OWNER	KRAFT Unbl.	KRAFT Bl.	SULFITE Unbl.	SULFITE Bl.	SODA Unbl.	SODA Bl.	NSSC Unbl.	NSSC Bl.	SOFT-WOOD %	HARD-WOOD %
			MISSISSIPPI (Continued)									
Moss Point	Jackson	International Paper Co. Sou. Kraft Div.	730	660							55	45
Natchez	Adams	International Paper Co. Sou. Kraft Div.	1000	1000							19	81
Vicksburg	Warren	International Paper Co. Sou. Kraft Div.	985	300							85	15
TOTAL MISSISSIPPI		Unbleached =	4405		0		0		0			
		Of Which Bleached =		1960								
		NUMBER OF MILLS	4			0		0		0		
			MONTANA									
Missoula	Missoula	Hoerner Waldorf Corp.	1200	250							100	0
TOTAL MONTANA		Unbleached =	1200		0		0		0			
		Of Which Bleached =		250								
		NUMBER OF MILLS	1			0		0		0		

TABLE A-2 (Continued)

CITY	COUNTY	OWNER	KRAFT Unbl.	KRAFT Bl.	SULFITE Unbl.	SULFITE Bl.	SODA Unbl.	SODA Bl.	NSSC Unbl.	NSSC Bl.	SOFT-WOOD %	HARD-WOOD %
NEW HAMPSHIRE												
Berlin	Coos	Brown Company	615	500							35	65
Berlin	Coos	Brown Company							220	0	0	100
Groveton	Coos	Groveton Papers Co.			115	105					100	0
Groveton	Coos	Groveton Papers Co.							225	0	0	100
Lincoln	Grafton	Franconia Paper Corp.			90	90					100	0
TOTAL NEW HAMPSHIRE		Unbleached =	615		205				445			
		Of Which Bleached =		500		195				0		
		NUMBER OF MILLS	1		2				2			
NEW YORK												
Deferiet	Jefferson	St. Regis Paper Co.			100	60					100	0
Glens Falls	Warren	Finch, Pruyn & Co., Inc.							237	178	0	100
Lyons Falls	Lewis	Georgia-Pacific Corp.							120	120	0	100
Mechanicville	Saratoga	West Virginia Pulp & Paper Co.							184	158	0	100
Norfolk	St. Lawrence	Northland Paper			60	0					100	0

T A B L E A-2 (Continued)

CITY	COUNTY	OWNER	KRAFT		SULFITE		SODA		NSSC		SOFT-WOOD	HARD-WOOD
			Unbl.	Bl.	Unbl.	Bl.	Unbl.	Bl.	Unbl.	Bl.	%	%
			NEW YORK (Continued)									
North Tonawanda	Niagara	International Paper Co.					155	155			0	100
Plattsburgh	Clinton	Georgia Pacific Corp.							70	0	0	100
Ticonderoga	Essex	International Paper Co.	205	190							0	100
TOTAL NEW YORK		Unbleached =	205		160		155		611			
		Of Which Bleached =		190		60		155		456		
		NUMBER OF MILLS	1		2		1		4			

T A B L E A-2 (Continued)

CITY	COUNTY	OWNER	KRAFT Unbl.	KRAFT Bl.	SULFITE Unbl.	SULFITE Bl.	SODA Unbl.	SODA Bl.	NSSC Unbl.	NSSC Bl.	SOFT-WOOD %	HARD-WOOD %
		NORTH CAROLINA										
Canton	Haywood	U.S. Plywood - Champion Papers, Inc.	1290	1290							60	40
Plymouth	Washington	Weyerhaeuser Co.	1180	450							80	20
Plymouth	Washington	Weyerhaeuser Co.							300	0	0	100
Riegelwood	Columbus	Riegel Paper Corp.	1070	1000							50	50
Roanoke Rapids	Halifax	Hoerner Waldorf Corp.	900	0							88	12
Sylva	Jackson	The Mead Corp.							270	0	0	100
TOTAL NORTH CAROLINA		Unbleached =	4440				0		570			
		Of Which Bleached =		2740						0		
		NUMBER OF MILLS	4		1			0	2			
		OHIO										
Chillicothe	Ross	The Mead Corp.	652	600							0	100
Circleville	Pickaway	Container Corp. of America							240	0	0	100
Coshocton	Coshocton	Stone Container Corp.							550	0	0	100
TOTAL OHIO		Unbleached =	652		0		0		790			
		Of Which Bleached =		600							0	
		NUMBER OF MILLS	1		0			0	2			

T A B L E A-2 (Continued)

CITY	COUNTY	OWNER	KRAFT Unbl.	KRAFT Bl.	SULFITE Unbl.	SULFITE Bl.	SODA Unbl.	SODA Bl.	NSSC Unbl.	NSSC Bl.	SOFT-WOOD %	HARD-WOOD %
						OKLAHOMA						
Broken Bow	McCurtain	Dierks Forests,Inc.							420	0	100	0
TOTAL OKLAHOMA		Unbleached =	0		0		0		420			
		Of Which Bleached =								0		
		NUMBER OF MILLS	0		0		0		1			
						OREGON						
Albany	Linn	Western Kraft Corp.	500	0							100	0
Coos Bay	Coos	Coos Head Timber Co.			90	0					100	0
Gardiner	Douglas	International Paper Co.	545	0							100	0
Jordan Cove	Coos	Menasha Corp.							250	0	0	100

T A B L E A-2 (Continued)

CITY	COUNTY	OWNER	KRAFT Unbl.	KRAFT Bl.	SULFITE Unbl.	SULFITE Bl.	SODA Unbl.	SODA Bl.	NSSC Unbl.	NSSC Bl.	SOFT-WOOD %	HARD-WOOD %
					OREGON (Continued)							
Lebanon	Linn	Crown Zellerbach Corp.			105	0					100	0
Newberg	Yarnhill	Publishers' Paper Co.			180	0					100	0
Oregon City	Clackamas	Publishers' Paper Co.			170	100					100	0
St. Helens	Columbia	Boise Cascade Corp.	800	650							100	0
Salem	Marion	Boise Cascade Corp.			220	210					100	0
Springfield	Lane	Weyerhaeuser Corp.	1150	0							100	0
Toledo	Lincoln	Georgia Pacific Corp.	950	0							100	0
Wauna	Clatsop	Crown Zellerbach Corp.	800	550							85	15
TOTAL OREGON		Unbleached =	4745		765		0		250			
		Of Which Bleached =		1200		310				0		
		NUMBER OF MILLS		6	5			0	1			

T A B L E A-2 (Continued)

CITY	COUNTY	OWNER	KRAFT Unbl.	KRAFT Bl.	SULFITE Unbl.	SULFITE Bl.	SODA Unbl.	SODA Bl.	NSSC Unbl.	NSSC Bl.	SOFT-WOOD %	HARD-WOOD %
		PENNSYLVANIA										
Erie	Erie	Hammermill Paper Co.							385	385	0	100
Johnsonburg	Elk	New York & Pennsylvania Co., Inc.	170	160							0	100
Johnsonburg	Elk	New York & Pennsylvania Co., Inc.			110	110					100	0
Lock Haven	Clinton	Hammermill Paper Co.					86	80			0	100
Mehoopany		Charmin Paper Products Co.			240	240					0	100
Roaring Springs	Blair	Combined Paper Mills, Inc.	195	180							50	50
Spring Grove	York	P. H. Glatfelter Co.	550	500							50	50
Tyrone	Blair	Westvaco	160	152							30	70
Williamsburg	Blair	Westvaco*	160	0							0	100
TOTAL PENNSYLVANIA		Unbleached =	1235		350		86		385			
		Of Which Bleached =		992		350		80		385		
			5		2		1		1			

*Shut down in 1969

T A B L E A-2 (Continued)

CITY	COUNTY	OWNER	KRAFT Unbl.	KRAFT Bl.	SULFITE Unbl.	SULFITE Bl.	SODA Unbl.	SODA Bl.	NSSC Unbl.	NSSC Bl.	SOFT-WOOD %	HARD-WOOD %
		SOUTH CAROLINA										
Catawba	York	Bowaters Carolina Corp.	860	860							88	12
Charleston	Charleston	Westvaco	1850	0							80	20
Florence	Florence	South Carolina Industries, Inc.	600	0							90	10
Georgetown	Georgetown	International Paper Co., Sou. Kraft Div.	1550	340							79	21
Georgetown	Georgetown	International Paper Co., Sou. Kraft Div.							325	0	0	100
Hartsville	Darlington	Sonoco Products Co.							400	0	0	100
TOTAL SOUTH CAROLINA		Unbleached =	4860		0		0		725			
		Of Which Bleached =		1200						0		
		NUMBER OF MILLS	4		0		0		2			

T A B L E A-2 (Continued)

CITY	COUNTY	OWNER		KRAFT Unbl.	KRAFT Bl.	SULFITE Unbl.	SULFITE Bl.	SODA Unbl.	SODA Bl.	NSSC Unbl.	NSSC Bl.	SOFT-WOOD %	HARD-WOOD %
		TENNESSEE											
Calhoun	McMinn	Bowaters Southern Paper Corp.		520	490							100	0
Counce	Hardin	Tennessee River Pulp & Paper Co.		700	0							91	9
Harriman	Roane	The Mead Corp.								190	0	0	100
Kingsport	Sullivan	The Mead Corp						275	260			0	100
Knoxville	Knox	Southern Extract Co.								119	0	0	100
TOTAL TENNESSEE		Unbleached	=	1220		0		275		309			
		Of Which Bleached	=		490				260		0		
		NUMBER OF MILLS		2			0	1		2			

T A B L E A-2 (Continued)

CITY	COUNTY	OWNER	KRAFT Unbl.	KRAFT Bl.	SULFITE Unbl.	SULFITE Bl.	SODA Unbl.	SODA Bl.	NSSC Unbl.	NSSC Bl.	SOFT-WOOD %	HARD-WOOD %
						TEXAS						
Evadale	Jasper	Eastex, Inc.	1200	1000							58	42
Houston	Harris	Southland Paper Mills, Inc.	450	100							100	0
Lufkin	Angelina	Southland Paper Mills, Inc.	325	300							100	0
Mulford	Orange	Owens-Illinois Inc.	1000	0							100	0
Pasadena	Harris	U. S. Plywood Champion Papers, Inc.	805	805							70	30
TOTAL TEXAS		Unbleached =	3780		0		0		0			
		Of Which Bleached =		2305								
		NUMBER OF MILLS	5		0		0		0			
						VIRGINIA						
Big Island	Bedford	Owens-Illinois, Inc.							450	0	0	100
Covington	Alleghany	Westvaco	950	850							34	66
Covington	Alleghany	Westvaco							270	0	0	100

T A B L E A-2 (Continued)

CITY	COUNTY	OWNER	KRAFT Unbl.	KRAFT Bl.	SULFITE Unbl.	SULFITE Bl.	SODA Unbl.	SODA Bl.	NSSC Unbl.	NSSC Bl.	SOFT-WOOD %	HARD-WOOD %
					VIRGINIA (Continued)							
Franklin	Isle of Wight	Union Camp Corp.	1100	940							43	57
Hopewell	Prince George	Continental Can Co., Inc.	900	0							100	0
Hopewell	Prince George	Continental Can							200	0	0	100
Lynchburg	Campbell	The Mead Corp.							190	0	0	100
West Point	King William	The Chesapeake Corp. of Virginia	1050	0							88	12
TOTAL VIRGINIA		Unbleached =	4000		0		0		1110			
		Of Which Bleached =		1790						0		
		NUMBER OF MILLS	4		0		0		4			
					WASHINGTON							
Anacortes	Skagit	Scott Paper Co.			135	135					5	95
Bellingham	Whatcom	Georgia Pacific Corp.			530	500					100	0

T A B L E A-2 (Continued)

CITY	COUNTY	OWNER	KRAFT Unbl.	KRAFT Bl.	SULFITE Unbl.	SULFITE Bl.	SODA Unbl.	SODA Bl.	NSSC Unbl.	NSSC Bl.	SOFT-WOOD %	HARD-WOOD %
		WASHINGTON (Continued)										
Camas	Clark	Crown Zellerbach Corp.	780	780							100	0
Camas	Clark	Crown Zellerbach			420	400					100	0
Cosmopolis	Grays Harbor	Weyerhaeuser Co.			488	400					90	10
Everett	Snohomish	Scott Paper Co.			850	850					100	0
Everett	Snohomish	Simpson Lee Paper Co.	85	80							0	100
Everett	Snohomish	Weyerhaeuser Co.	440	400							85	15
Everett	Snohomish	Weyerhaeuser Co.			360	310					100	0
Hoquiam	Grays Harbor	ITT Rayonier,Inc.			545	545					100	0
Longview	Cowlitz	Longview Fibre Co.	1800	400							100	0
Longview	Cowlitz	Longview Fibre Co.							140	0	0	100
Longview	Cowlitz	Weyerhaeuser Co.	695	650							91	9
Longview	Cowlitz	Weyerhaeuser Co.			425	400					100	0
Longview	Cowlitz	Weyerhaeuser Co.							230	0	100	0
Millwood	Spokane	Inland Empire			42	20					100	0
Port Angeles	Clallam	Fibreboard Corp.			70	65					95	5
Port Angeles	Clallam	ITT Rayonier, Inc.			450	450					100	0

T A B L E A-2 (Continued)

CITY	COUNTY	OWNER	KRAFT		SULFITE		SODA		NSSC		SOFT-WOOD	HARD-WOOD
			Unbl.	Bl.	Unbl.	Bl.	Unbl.	Bl.	Unbl.	Bl.	%	%
			WASHINGTON (Continued)									
Port Townsend	Jefferson	Crown Zellerbach Corp.	420	0							100	0
Tacoma	Pierce	St. Regis Paper Co.	930	145							100	0
Vancouver	Clarke	Boise Cascade Corp.			200	185					100	0
Wallula	Walla Walla	Boise Cascade Corp.	410	0							100	0
Wallula	Walla Walla	Boise Cascade Corp.							140	0	100	0
TOTAL WASHINGTON		Unbleached =	5560		4515		1		510			
		Of Which Bleached =		2375		4260				0		
		NUMBER OF MILLS	8			12		0	3			

T A B L E A-2 (Continued)

CITY	COUNTY	OWNER	KRAFT Unbl.	KRAFT Bl.	SULFITE Unbl.	SULFITE Bl.	SODA Unbl.	SODA Bl.	NSSC Unbl.	NSSC Bl.	SOFT-WOOD %	HARD-WOOD %
WISCONSIN												
Appleton	Outagamie	Consolidated Papers, Inc.			165	155					100	0
Brokaw	Marathos	Wausau Paper Mills Co.			155	145					15	85
Green Bay	Brown	American Can Co.			155	145					0	100
Green Bay	Brown	Charmin Paper Products Co.			130	123					100	0
Green Bay	Brown	Green Bay Packaging, Inc.							250	0	0	100
Kaukauna	Outagamie	Thilmany Pulp & Paper Co.	380	90							100	0
Marinette	Marinette	Scott Paper Co.			55	50					100	0
Mosinee	Marathon	Mosinee Paper Mills Co.	175	0							80	20
Nekoosa	Wood	Nekoosa-Edwards Paper Co.	340	315							75	25
Oconto Falls	Oconto	Scott Paper Co			115	110					80	20
Park Falls	Price	Kansas City Star Co.			115	110					67	33
Peshtigo	Marinette	Badger Paper Mills, Inc.			100	90					75	25
Port Edwards	Wood	Nekoosa-Edwards Paper Co.			235	225					0	100

T A B L E A-2 (Continued)

CITY	COUNTY	OWNER	KRAFT Unbl.	KRAFT Bl.	SULFITE Unbl.	SULFITE Bl.	SODA Unbl.	SODA Bl.	NSSC Unbl.	NSSC Bl.	SOFT-WOOD %	HARD-WOOD %
WISCONSIN (Continued)												
Rhinelander	Oneida	St. Regis Paper Co.			217	200					100	0
Rothschild	Marathon	American Can Co.			217	200					0	100
Tomahawk	Lincoln	Owens-Illinois Inc.							615	0	0	100
Wisconsin Rapids	Wood	Consolidated Papers, Inc.	360	333					67		33	
TOTAL WISCONSIN		Unbleached =	1255		1552		0		865			
		Of Which Bleached =		738		1463				0		
		NUMBER OF MILLS	4		11			0	2			

TABLE A-3

ANNUAL PRODUCTION AND CONSUMPTION OF CHEMICAL WOOD PULPS IN U.S.A. 1937 - 1967

MILLIONS OF TONS OF PULP

Year	Kraft	Sulfite	NSSC	Dissolving	Soda	Total Production Chemical Pulps
1937	2.1	1.8	0.1	0.4	0.5	4.9
1940	3.7	2.3	0.2	0.3	0.5	7.0
1945	-	-	-	-	-	9.3
1950	7.5	2.4	0.7	0.5	0.5	11.6
1953	9.4	2.3	1.0	0.7	0.4	13.8
1955	11.3	2.6	1.4	1.0	0.4	16.7
1956	12.1	2.7	1.5	0.9	0.5	17.7
1957	11.9	2.6	1.6	1.0	0.4	17.5
1958	12.3	2.4	1.6	0.9	0.4	17.6
1959	13.8	2.5	1.9	1.1	0.5	19.8
1960	14.6	2.6	2.0	1.1	0.5	20.7
1961	15.4	2.6	2.4	1.2	0.4	22.0
1962	16.3	2.6	2.5	1.3	0.4	23.1
1963	17.5	2.7	2.6	1.4	0.4	24.6
1964	20.1	2.7	2.7	1.5	0.2	27.2
1965	21.5	2.7	2.9	1.5	0.2	28.8
1966	22.4	2.8	3.2	1.5	0.2	30.1
1967	22.8	2.7	3.3	1.4	0.2	30.4

Year	Net Imports	Consumption Chem. Pulps	Consumption Per Capita - Pounds
1937	1.9	6.8	105
1940	0.6	7.6	115
1945	4.1	10.4	151
1950	2.0	13.6	180
1953	1.7	15.5	194
1955	1.3	18.0	218
1956	1.4	19.1	226
1957	1.2	18.7	210
1958	1.4	19.0	216
1959	1.6	21.4	240
1960	1.0	21.7	240
1961	1.0	23.0	250
1962	1.3	24.4	261
1963	1.0	25.6	271
1964	1.0	28.2	295
1965	1.4	30.2	310
1966	1.5	31.6	320
1967	1.2	31.6	318

Source: U. S. Bureau of Census
U. S. Pulp Producers Association
American Paper Institute

TABLE A-4

CHEMICAL WOOD PULP MILLS
SUPPLEMENTARY DATA
CURRENT AND PLANNED NEW PLANT CONSTRUCTION
AS OF DECEMBER 31, 1968

State	Owner	New - N Exp.- E	Kind of Pulp & Capacity - AD Tons/Day			Date on Stream
			Kraft	Sulfite	NSSC	
Alabama	U.S. Plywood-Champion Papers	N	500 Bl.			1970
Alaska	U.S. Plywood-Champion Papers	N	600			1973
Arkansas	Georgia-Pacific Corp.	E	425 Bl.			1970
	Potlatch Forests, Inc.	N	400 Bl.			-
Florida	Container Corp. of America	E	550		350	1971
Idaho	Potlatch Forests, Inc.	E	220			1969
Kentucky	Westvaco	N	600 Bl.			1970
	Western Kraft Corp.	N	200 Bl.			1969
Louisiana	Boise Cascade Corp.	N	600			1970
Maine	Great Northern Paper Co.	E		480		1969
Minnesota	Northwest Paper Co.	E	250 Bl.			1971
New York	Finch, Pruyn & Co., Inc.	N	510 Bl.			1970
	International Paper Co.	N	510 Bl.			1970

TABLE A-4 (Continued)

North Carolina	Georgia-Pacific Corp.*	N				200	1969
	Weyerhaeuser Corp	N		600 Bl.			1970
Oregon	American Can Company	N		300 Bl.			1969
Pennsylvania	Hammermill Paper Co.	E				200 Bl.	1970
South Carolina	Kimberly-Clark Corp.	N		300 Bl.			-
Tennessee	Inland Container Corp.	N				300	1970
Texas	International Paper Co.	N	650	650 Bl.			1971
Virginia	Union Camp Corp.	E		600 Bl.			1970
Washington	ITT Rayonier, Inc.	N		300 Bl.			-
West Virginia	Milburn Corp.	N				250	-
TOTAL - AD TONS PER DAY				7,605	830	1,300 =	9,735 Grand Total
TOTAL - AD TONS PER YEAR - 350 DAYS				2,662,000	290,000	455,000 =	3,407,000

* Plans shelved in 1969

Source: Paper Trade Journal - February 3, 1969
Pulp & Paper - December 16, 1969

TABLE A-5

CHEMICAL WOOD PULP MILLS
SUPPLEMENTARY DATA
PROPOSED AND TENTATIVE NEW MILLS
AS OF DECEMBER 31, 1968

Location	Owner	New - N Exp.- E	TPD	Kind	Stage
South Carolina	Tenneco Co.	N	-	Kraft	Study
Wyoming	State	N	500	Bl. Kraft	Study
Florida	Great Northern Paper Co.	N	300	Bl. Kraft	Study
Minnesota	Boise Cascade Corp.	N	500	Kraft	Study
Mississippi	Columbia Pulp & Paper Co.	N	350	Kraft	Deferred
Minnesota	Minnesota International Pulp & Paper Co.	N	-	Newsprint	Study
Arizona	Ponderosa Paper Products, Inc.	N	-	Sulfite	Planned
Texas	Sabine Pulp & Paper Co.	N	-	Bl. Kraft	Planned
Texas	Texas Newsprint Co.	N	300	Newsprint	Study
Missouri	Delta-New Madrid Paper Corp.	N	-	Newsprint	Study
Oklahoma	Dierks Forests, Inc.	N	-	-	Proposed
Michigan	Oxford Paper Co.	N	400	Bl. Kraft	Deferred

Source: Paper Trade Journal
Pulp & Paper

TABLE A-6

CHEMICAL WOOD PULP MILLS
SUPPLEMENTARY DATA

ESTIMATE OF PHASED OUT OPERATIONS
1968 THROUGH 1970

MILLS CLOSED DOWN

In 1968 - (Tonnages not included in 1968 Mill Capacities)

Location			Sulfite	Soda	Kraft	NSSC
Maine	May	1968	125			
Minnesota	Early	1968	180			
Oregon	May	1968	210			
Wisconsin	June	1968	130			
New York		1968		60		
Wisconsin		1968	190	(Tonnage replaced by new kraft mill)		
	Total		835	60	0	0
AD Tons per	350 Day Year		292,000	21,000	0	0

In 1969 and 1970 (Tonnages included in 1968 Mill Capacities)

Michigan	May	1969	168			
New York		(1970)	(To be replaced by New Mill)		205	
Minnesota		1970	135			
Washington	Feb.	1969	200			
New York		1969		(To be replaced by MgO)		235
	Total		503	0	205	235
Ad Tons per	350 Day Year		176,000	0	72,000	82,000

Maine - In 1969 - A 500 TPD Sulfite Sodium Base Mill will be converted to Magnesium Base.

Note: Probably other Sulfite Mills using Sodium and Calcium Base Processes will convert to other Processes, but announcements have not been made up to this time.

Source: This study (Meakin)

TABLE A-7

CHEMICAL WOOD PULP MILLS
SUPPLEMENTARY DATA

ESTIMATE OF PHASED OUT OPERATIONS
1970 TO 1980

State	Number Mills	Process and AD Tons per Day			
		Kraft	Sulfite	Soda	NSSC
Massachusetts	1			54	
New Hampshire	2		205		
New York	2		160		
Oregon	2		195		
Pennsylvania	2		110	86	
Washington	3	85	112		
Wisconsin	7		780		
Total	19	85	1,562	140	0
AD Tons per 350 Day Year		30,000	547,000	49,000	